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This document is a compilation of 68 papers presented at the First National Symposium/Workshop on Rotating Biological Contactor (RBC) Technology sponsored by the University of Pittsburgh in cooperation with the US Army Construction Engineering Research Laboratory (Champaign, IL) and the USEPA Office of Research and Development's Municipal Environmental Research Laboratory (Cincinnati, Ohio). The Symposium/Workshop was held February 4-6, 1980 at Champion, Pennsylvania. The papers presented in the three-day Symposium and the findings of the research needs Workshop comprise the major portion of these proceedings.

The Symposium/Workshop proceedings will document present knowledge regarding RBC technology. The papers are divided into 11 major topic areas:

1. Perspective, Overview, History
2. Process Variables and Biofilm Properties
3. Municipal Wastewater Treatment
4. Biokinetic Studies
5. Air Drive and Supplemental Aeration
6. Industrial Wastewater Treatment
7. Concepts and Models
8. Upgrading Primary and Secondary Waste Treatment Systems with RBC's
9. Design and Operation
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THE BIOLOGICAL APPROACH TO THE ROTATING DISC PROCESS

By

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President
Aquamotion, Inc.
Minneapolis, Minnesota

INTRODUCTION

Much of our (United States of America) current water and wastewater treatment technology has been imported from Europe. The Rotating Biological Surface (RBS) process was also conceived and thereafter commercialized in Europe. Professors Hartmann and Popel deserve the majority of the research credit for the process as we know it today. A short twenty five years ago, in Stuttgart, West Germany, they were rotating discs and were most careful in observing and recording the several phenomena which had occurred throughout their test programs. In 1965, after ten years of intensive RBS process development work, Dr. Hartmann[1] cautioned against loading the first stage of an RBS process multistage system beyond about 20 lbs BOD/1000 sq ft/day. This biological caveat, although followed throughout the common market, was cast aside by the RBS process practitioners in this country. This paper attempts to explain where we went wrong, and what we must now do in order to optimize RBS process applications.

Overcoming and thereafter modifying the existing gross misunderstandings of the RBS process will not be quick or easy. Wastewater treatment technology changes most slowly. Lag time between process design and actual plant operation is several years. It is hoped that the disclosures herein will assist in the diagnosis of RBS process problems and their cure. By applying the organic loading design principles enunciated by Hartmann, future RBS process problems ought to be minimized, if not eradicated.

BACKGROUND

The first West German installations of the RBS process were on rather small sanitary waste applications. Disc plants were designed on the basis of number of inhabitants. Since treatment plant designers were well aware of the correlation between number of people and volume of liquid wastes, and while not unmindful of the 20 lbs/1000 sq ft maximum organic loading/day limitation, a "short hand" hydraulic design basis was developed which utilized gpm/sq ft disc area/day. As combinations of sanitary/industrial wastes were gradually encountered, the hydraulic approach was modified to "population equivalents" so as to take into consideration the organic loading of the industrial waste contribution. As a result of this consistent observation of and respect for biology, RBS process failures in the common market over the last 20 years are virtually non-existent.

In this country, the RBS process was boldly and confidently applied frequently when biological treatment was possible. By relying upon the supplier sponsored pilot testing and thereafter adjusting the hydraulic design approach, RBS process designs were formulated on high strength industrial wastes and nitrification applications. However laudable the effort to expand the frontiers of this technology, to do so beyond the scientific method was tantamount to driving beyond the stopping range of a car's headlights. As a result, several RBS process crashes have occurred and more are imminent. We must now return to the realities of biology while discarding the convention of hydraulics in order to properly understand the RBS process.

DESIGN CONSIDERATIONS

Some [2,3] investigators have discussed the importance of organic loading vis-a-vis hydraulic turbulence, disc rpm, dissolved oxygen, oxygen mass transfer, and associated considerations. That organic loading is the single and exclusive basis for RBS process design was first disclosed by Steiner.[4] The hydraulic design approach has lost its credibility. Indeed, the express purpose of this Symposium is to discover the true bases upon which the RBS process is dependent, and, if thought advisable, to recommend additional RBS process testing.

By plotting organic loading versus organic removal, a typical 4-stage (with equal surface area per stage) RBS process system would appear as shown in Figure 1.

RBS process equipment suppliers,[5,6,7,8,9,10,11,12] however, publish design curves of two types, namely those characterized by Figures 2 and 3.

In calculating the mass organic loading versus BOD removal represented by either family of curves in Figures 2 and 3, we again obtain a curve identical to Figure 1! See Figure 4.

In fact, the characteristic curve shown in Figure 4 is consistent with all RBS process data published worldwide to date. The conclusion, therefore, that it truly represents the RBS process becomes virtually inescapable. The extreme simplicity of the RBS process is based upon this curve and the fact that the slope and extension of the curve are its only variables.

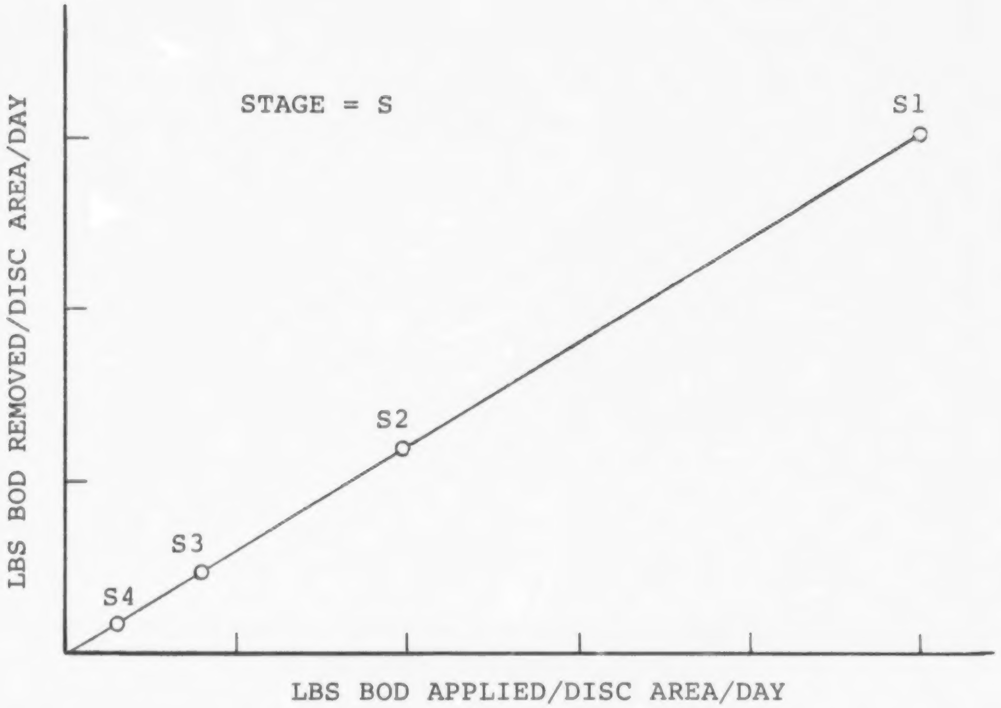


FIGURE 1. Typical four-stage RBS process treatment curve.

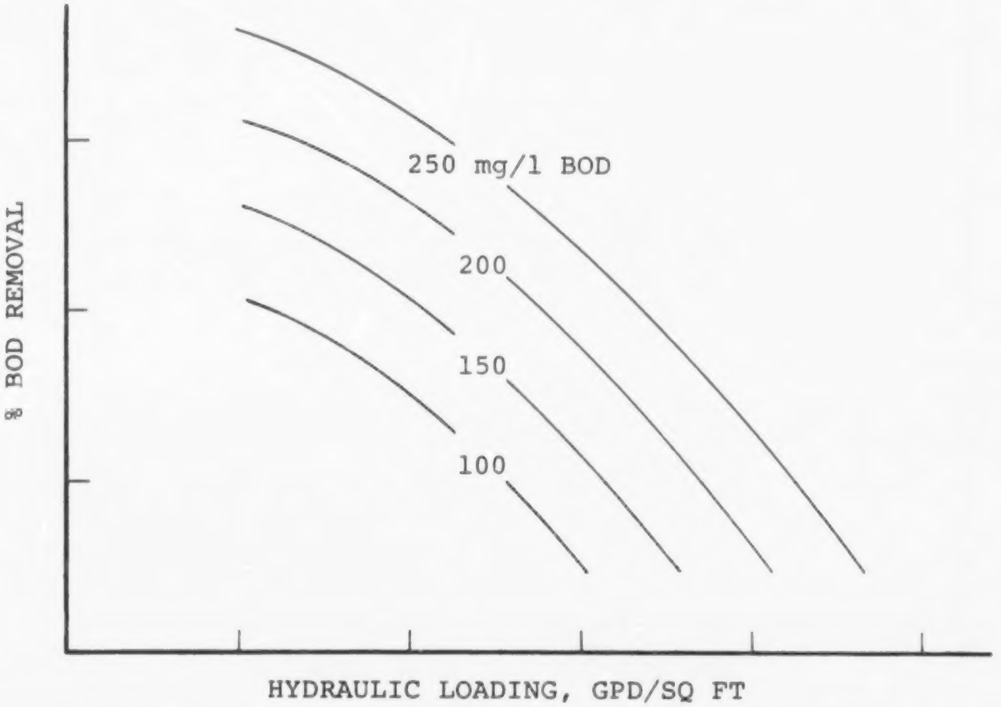


FIGURE 2. Hydraulic loading versus percent BOD removal.

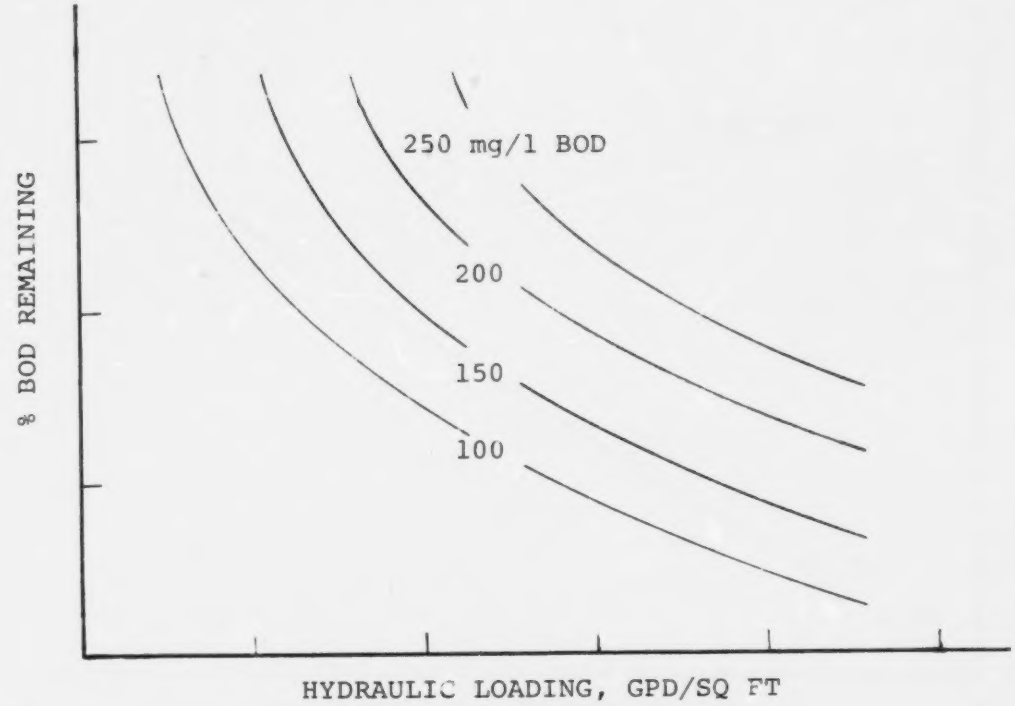


FIGURE 3. Hydraulic loading versus percent BOD remaining.

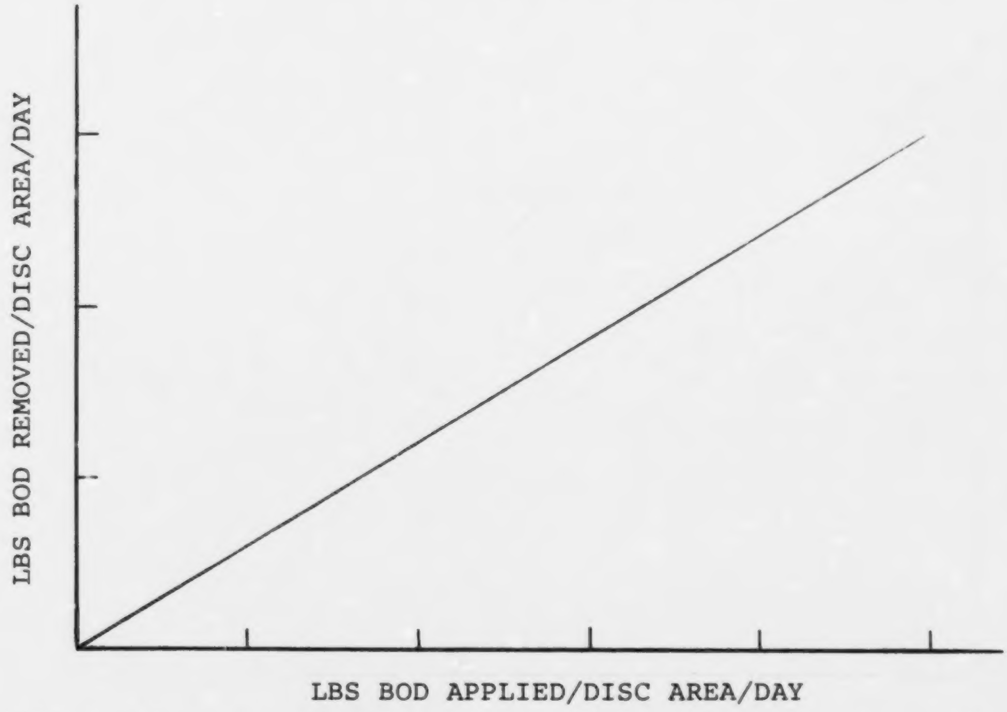


FIGURE 4. Characteristic RBS process treatment curve obtained from RBS process equipment design manuals and trade literature.

The slope of the curve is exclusively influenced by the treatability of the waste. Both cooler temperatures and the presence of inhibitory toxins retard microbiological activity and the treatment curve lessens. However, a full 50% decline in treatability in a typical four-stage system would result in only a theoretical 27.1% decline in effluent quality.[4]

The extension of the curve is limited by either oxygen mass transfer into the liquid waste, or limiting microbiological activity, whichever first occurs. In BOD removal systems, disc rpm control the removals obtainable, while in nitrification applications, metabolic rates control.

The significance of this RBS process curve disclosure becomes evident as we again examine the treatment characterized in Figure 1. Lets attach some typical removal kinetics to the graph while progressively reducing the square feet of surface area by 1/2 per stage. That is, stage 2 now has 1/2 the disc surface area of stage 1, stage 3 has 1/2 of stage 2 or 1/4 of stage 1, and stage 4 has 1/2 of stage 3, or 1/4 of stage 2, or 1/8 of stage 1. The new loadings and removals (compared with Figure 1) are shown by the revised points in Figure 5.

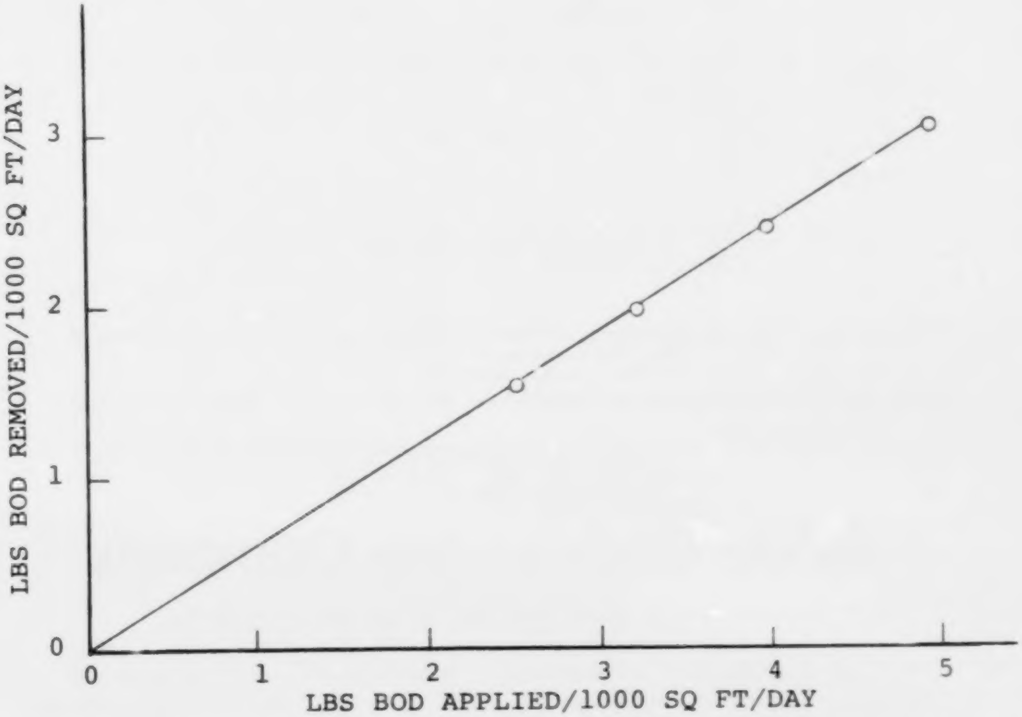
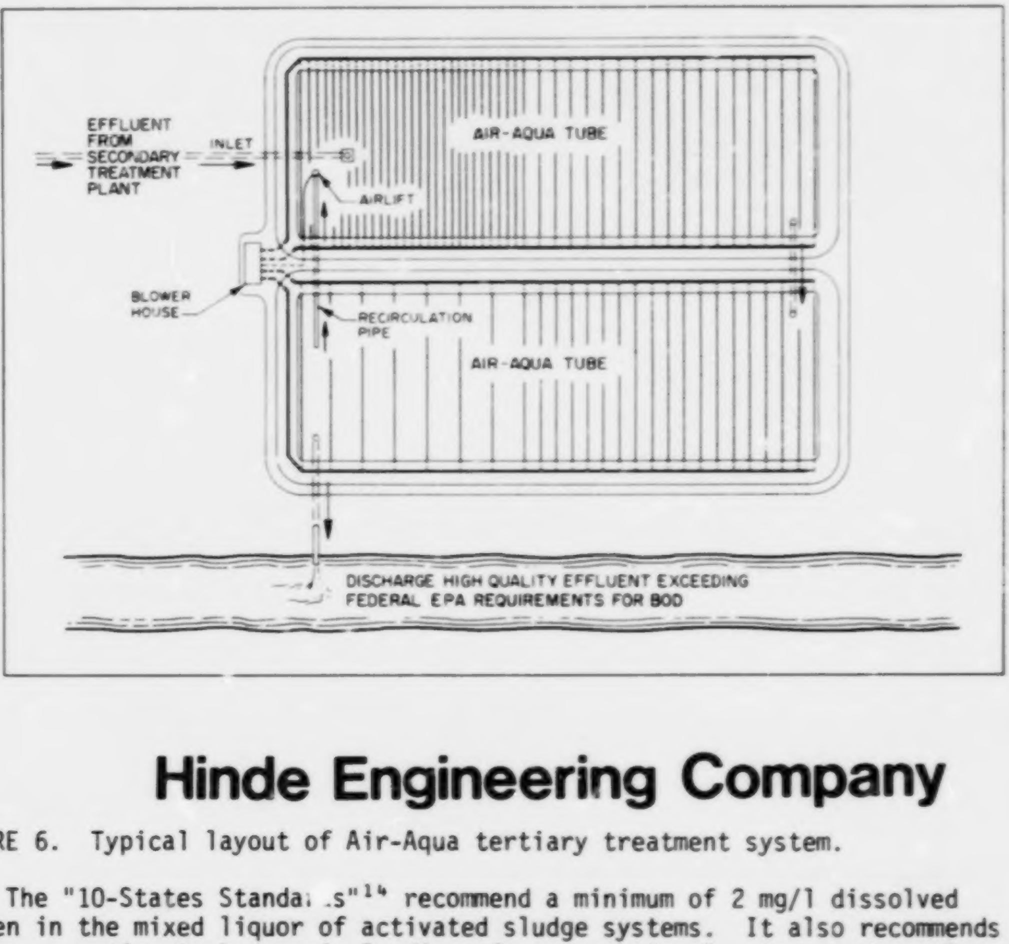


FIGURE 5. Typical RBS process treatment curve of 60% removal per stage.

In now adding up the mass removals per stage, we arrive at the same number of lbs BOD removed as with the equal surface area per stage system represented in Figure 1. The Figure 5 system, however, uses less than 1/2 the disc surface area used in Figure 1 (and therefore installed and energy costs are also halved).

DISCUSSION

From a microbiological purview, there is absolutely nothing profound with the RBS process. The literature, for example, is filled with similar references to the simultaneous increase in dissolved oxygen and the concurrent decrease in biological growth in each successive treatment stage. This phenomenon is nothing more than a restatement that aeration requirements decrease with each increase in treatment stage. With Hinde's[13] diffused air aeration system, their recommendation as shown in Figure 6 calls for successive less aeration as more treatment is achieved. Note the progressive decrease in aeration tubing concurrent with increase in treatment.


EFFLUENT FROM SECONDARY TREATMENT PLANT
INLET
AIR-AQUA TUBE
AIRLIFT
BLOWER HOUSE
RECIRCULATION PIPE
AIR-AQUA TUBE
DISCHARGE HIGH QUALITY EFFLUENT EXCEEDING FEDERAL EPA REQUIREMENTS FOR BOD

Hinde Engineering Company

FIGURE 6. Typical layout of Air-Aqua tertiary treatment system.

The "10-States Standards"[14] recommend a minimum of 2 mg/l dissolved oxygen in the mixed liquor of activated sludge systems. It also recommends maximum aeration tank organic loadings for conventional, step, and complete mix aeration; contact stabilization; and extended aeration and oxidation ditch methods of treatment. Hydraulics seem to be completely lacking in the recognized design manuals for biological systems other than the RBS process. Absent therefore is the requirement for equal aeration in successive stages of RBS process treatment which is represented by equal surface areas per stage. And since the later stages of disc surfaces can physically support a biological population equal in density to stage 1, there remains no reason to have equal surface areas per successive treatment stage, since zoogloea slimes neither know nor care what particular stage of treatment they inhabit.

FIELD PROBLEMS

The Spencer, Iowa RBS process facility was designed for carbonaceous BOD removal followed by RBS process nitrification. It is a ten-stage system with 100,000 sq ft of surface area in the first six stages and 150,000 sq ft of surface area in the last four stages. Plant flow is equally divided into four flow streams through the RBS process facility. A local meat packing plant was to contribute about 75% of the organic loading with domestic wastes accounting for the balance. During the first year of RBS process plant operation, the meat packing plant was idle due to a strike. The then treatment curve is approximately represented as shown in Figure 7.

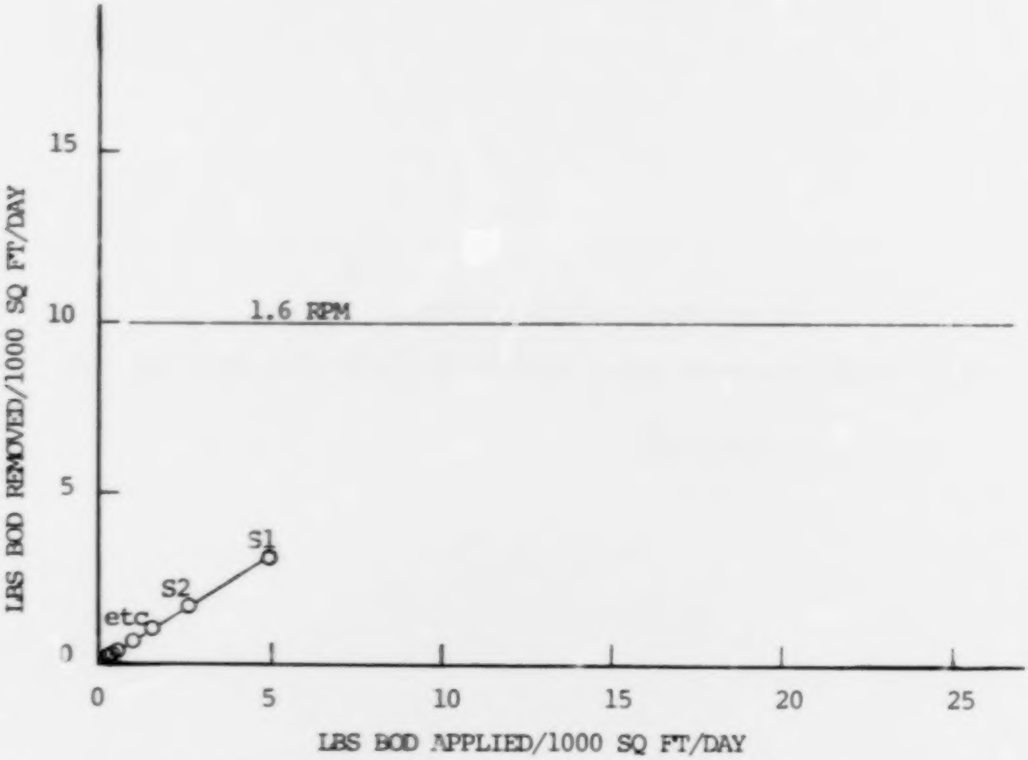


FIGURE 7. RBS process treatment curve for Spencer, Iowa without meat packing wastes.

The packing plant was eventually purchased by Land O'Lakes and as their production capacity gradually increased to pre-strike levels, the RBS process treatment plant began going anaerobic. Treatment is now represented by the curve in Figure 8. Since the entire organic load is passed through stage 1, and because the mechanical aeration capability of stage 1 is fixed by its rpm (and therefore limited), the metabolic oxygen demand goes unsatisfied. As a result, the first stages have already gone anaerobic and its now merely a matter of time before the effluent quality deteriorates beyond design expectations, with concurrent violations of Iowa DEQ discharge regulations.

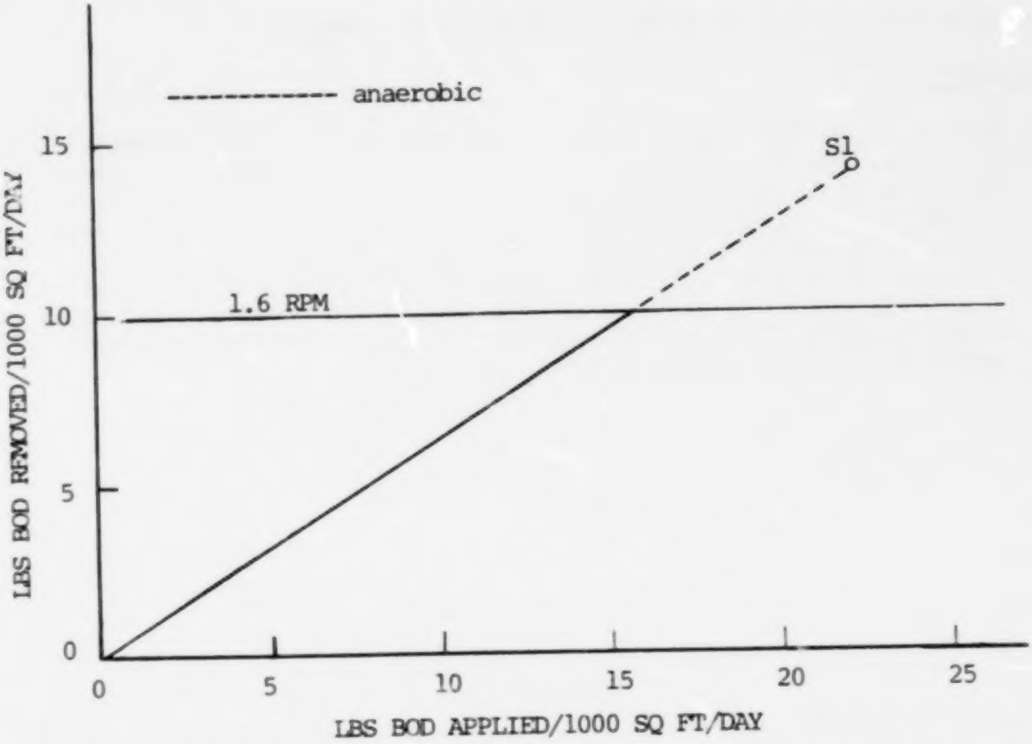


FIGURE 8. RBS process treatment curve for Spencer, Iowa with meat packing wastes.

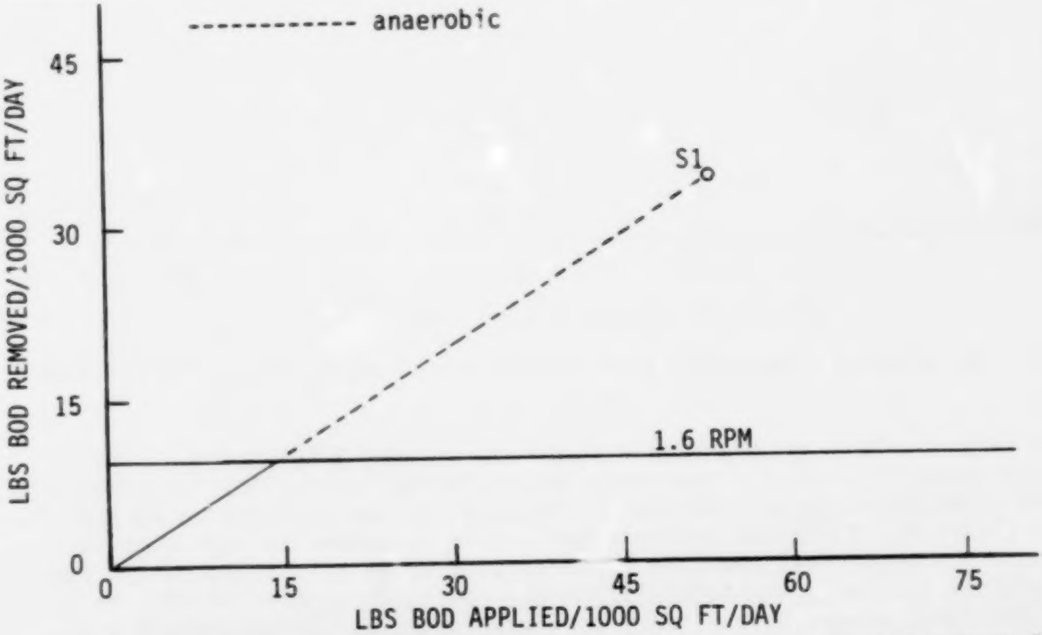


FIGURE 9. RBS process treatment curve for filter cake pressate return liquor at Minneapolis/St. Paul Metropolitan Waste Control Facilities.

Figure 9 represents the design curve for a waste which is extremely high (4,000 to 5,000 mg/l BOD) in organic strength. At the time of preparing this article, the Twin Cities Metropolitan plant had yet to start. Its failure is sure to occur when evaluated on the organic loading design principles.

Generally, when carbonaceous BOD and Nitrification are designed for in the same flow stream using multiple stages, or when high strength organic wastes are subject to multistage treatment based on hydraulics, RBS process problems are likely to occur. Figures 8 and 9 represent dozens of existing like designs. Resulting RBS process failures should be attributable to the hydraulic design approach rather than the viability of the process itself.

NITRIFICATION

The organic loading versus removal curve which represents the carbonaceous kinetics also represents the RBS Nitrification process. When one plots data on loadings versus removals,[15,16,17] a similar curve is produced, except that the slope is far less. See Figure 10.

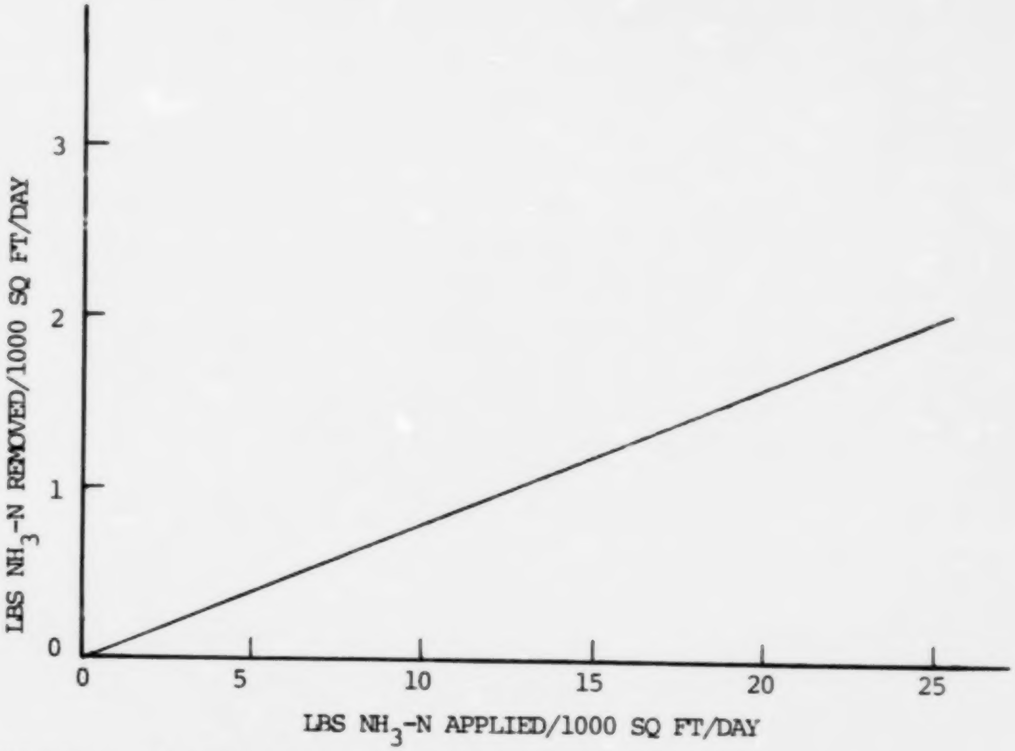


FIGURE 10. Typical RBS process nitrification curve.

It follows from the above curve that high NH_3-N loadings/surface area coupled with multiple stages will each enhance removals while reducing RBS process equipment requirements.

CONCLUSIONS

Since the RBS process is immune from the hydraulic considerations of residence time and hydraulic loading, it may be considered for the treatment of:

1. Combined sewer overflows

2. Inflows and infiltration

For full scale proof of this, the Spencer, Iowa plant is connected to a combined sewer system and has withstood scores of hydraulic surges during the last year and one half without impairing treatment.

RECOMMENDATIONS

The organic design approach will make the RBS process at least twice as attractive from installed costs and energy consumption considerations when viewed from current state-of-the-art. Comprehensive RBS process testing should be initiated to prove the organic loading approach respecting both carbonaceous and nitrification removals.

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FACTORS AFFECTING ATTACHMENT AND DEVELOPMENT OF BIOLOGICAL FILMS ON SOLID MEDIA

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INTRODUCTION

Attachment of biological films on solid surfaces is of considerable importance to engineers. Biological film development may constitute a nuisance in conveyance lines, heat exchangers and cooling towers. In cooling systems, biofilms increase frictional resistances, accelerate corrosion and impair heat transfer. Alternately, sanitary engineers have used biofilms in wastewater treatment plants for BOD and nitrogen removal.

Although there is some knowledge about the dynamics of film development, and a considerable body of information about the kinetics of substrate removal by biofilms, the mechanisms and factors governing film attachment are still unclear.

Work done by marine biologists and engineers has shown that colonization of a solid surface by bacteria is a rather selective process, which is affected by environmental conditions. Ionic strength, pH, substrate and nutrient concentration, water temperature, the surface properties of the solid, are all thought to influence biofilm attachment and development.

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Despite the obvious importance of biological films, there is paucity of information on the dynamics of biofilm development in the sanitary engineering literature. This paper summarizes a good deal of the available information on the subject, and presents the results of preliminary research on the effect of surface preconditioning on the formation and growth of biological films.

BACTERIAL ATTACHMENT

Zobell and Anderson(1) were among the first researchers to explore the phenomenon of bacterial attachment. They observed that bacterial activity in seawater samples stored in small containers was significantly higher than the activity in samples stored in larger containers of the same shape. This was attributed to the proportionately larger amount of surface area provided by the small containers, which served to concentrate nutrients from dilute media such as seawater.

Enhnaced bacterial activity in the presence of submerged surfaces was also observed by Heukelekian and Heller(2). Using unpolluted river water, they studied the effect of different types of surfaces and agitation on bacterial growth. The samples to which sand was added consistently had much higher bacterial concentrations than those samples that were aerated, agitated with paddles or left quiescent.

In later studies Zobell(3) determined that growth enhancement by surfaces was more pronounced at low nutrient concentrations. The distribution of the submerged surfaces was also found to affect the efficiency of growth enhancement. In the same study, Zobell suggested that there is a relationship between solubility, dispersion or molecular size and the nutrient concentration phenomenon. Increasing the surface to volume ratio apparently had little effect in dilute solutions of glucose, glycerol or lactate, but did affect solutions enriched with sodium caseinate, lignoprotein or an emulsified chitin preparation.

Past research has demonstrated that bacterial attachment is not a simple consequence of increased surface to volume ratio. The factors affecting attachment will be discussed in later sections of this paper. First, it is of interest to discuss the steps occurring during the process of bacterial attachment.

The Attachment Process

Development of biological films on solid surfaces is the result of several processes. Characklis and co-workers(4,5) have summarized the steps contributing to the overall biofilm development as follows: adsorption of organics at the wetted surface, transport of microorganisms from the bulk of the liquid to the wetted surface, microorganism attachment to the solid surface, metabolism and growth of the attached microorganisms (biofilm growth), and detachment or re-entrainment of biofilm due to fluid shear stress. The present work will focus on the mechanisms of bacterial attachment, and on the factors which influence the irreversible adhesion of bacteria on solid surfaces.

Marshall and co-workers(6,7,8,9) have proposed that the attachment process involves two distinct phases, each controlled by different mechanisms.

The first phase is considered to be an almost instantaneous phase, where bacterial adhesion is relatively week and reversible. The second phase is one of irrversible firm attachment, requiring an "incubation period", described earlier by Zobell(1), of at least 3 hours. This lag period is presumably the time required for the bacteria to produce the exocellular "cements" necessary to achieve firm attachment. Hendricks(10), using river water, observed the two phases proposed by Marshall for heterotrophic and enteric bacteria. After an initial rapid rate of attachment that was equivalent to a doubling time of 2 hours, the rate of adsorption of bacteria onto glass slides slowed down, plateauing at about 24 hours.

Work done by many researchers over the past decade strongly indicates that an acidic polymer produced by the bacteria is in some way integrally involved in phase-2 attachment. It should be noted that a few bacteria appear to attach to surfaces via specialized appendages or holdfast structures. These bacteria are rather unique however, and do not reflect the normal mode of bacterial attachment. They will not be considered further in this paper.

Much was known in the fields of colloid and surface sciences by the time researchers began to study bacterial attachment. Quite naturally investigators have attempted to model biological adhesion using analogies to existing theories, much as Stumm and co-workers(11,12) and Pavoni et al.(36) did in describing bacterial aggregation and flocculation. While these theories have proved useful in extending the existing knowledge of bioadhesion, the complexity of the process, which includes biological reactions, has prevented the formulation of a unified theory that satisfactorily describes all situations.

Both phases of the attachment process will be looked at in greater detail in the following sections.

Phase 1: Reversible Adsorption

The initial adsorption of bacteria to submerged surfaces occurs almost instantaneously(3,6,13). Fletcher(14) modeled the rate of attachment of marine pseudomonad using a modified Langmuir adsorption isotherm. This adsorption, as reported by Daniels(15), ceases for all practical purposes after 15 minutes.

During this first phase of attachment both motile and non-motile bacteria remain in motion. Motile bacteria have been found to rotate violently around their axis, sometimes breaking loose from their adsorption site(15), while non-motile bacteria exhibit Brownian motion(6). This has led some researchers to believe that there is no true adsorption of bacteria on the surface in this first phase. Marshall(6) expanded the Derjaguin-Landau-Vervey-Overbeck model, which Weiss(16) had earlier used for animal cells, and postulated that the bacteria are held at a small but finite distance from the surface by physical forces, which result from a balance of the van der Waals attractive forces with the electron double layer replusive forces.

Research by Meadows(17,18) confirmed the observations made by Marshall with marine bacteria. In addition, this researcher found that optimum attachment for marine and freshwater bacteria occur at radically different ionic strengths.

The factors affecting the reversible adsorption of microorganisms have been summarized by Daniels(15), and are presented in Table 1. Fletcher(14) studied the effect of culture age, time, temperature and growth phase on phase-1 adsorption. Cells were found to adsorb best in log growth and worst in the death phase. The time required to reach steady adsorption onto polystyrene was 2.5-3.0 hours for log growth, 3.4-4.0 hours for the stationary phase and greater than 5.0 hours for the death phase. The number of cells attaching was found to be proportional to the bacterial concentration in solution until the surface became saturated. A Langmuir isotherm fit the data well. As the temperature was decreased, the rate of adsorption was also reduced, but given sufficient time the steady state number of cells adsorbed was the same. Thirty-five cells per 100 μm^2 appeared to be the adsorptive limit.

Changes in pH can either cause desorption of previously adsorbed cells, or adsorption of cells from solution. This has been explained by Daniels(15) as a charge reversal phenomenon of specific chemical groups at the cell surface. Adsorption occurs when the resulting charge at the cell wall and that of the submerged surface are opposite. The electrostatic forces which promote cell adhesion, in this case, are in general an order of magnitude larger than the physical adsorptive forces described by Marshall(6).

Adsorption between two negatively charged surfaces can be made possible by cation bridging. Daniels(15) reports that addition of divalent cations to the growth media promotes attachment between bacteria and the solid surface.

Regardless of the type of adsorption, namely by van der Waals forces or by coulombic attraction, the concensus is that the initial phase of bacterial attachment does not result in a permanent bond; in fact, in many cases, it can be readily removed by even mild rinsing. The type of bond required to withstand shear forces, such as those prevalent in engineering systems (pipelines, rotating biodiscs, fluidized beds), can only be obtained by means of a bonding agent between the cells and the solid surface. This "cement", which is excreted by bacteria, produces permanent attachment, and allows growth of a healthy and stable biological film.

Phase 2: Permanent Bacterial Attachment

Bacteria can stick to surfaces with tenacity. Zobell(1,3) observed that repeated rinses with running water were insufficient to dislodge bacteria growing attached to a solid surface. A few researchers have attempted to estimate the adhesive strength between the surface and the bacteria. Zvyaginsten(19) determined the adhesive force of several gram-negative and gram-positive bacteria via a centrifugal rputure technique. Adhesive forces ranged from 4×10^{-7} to 4×10^{-4} dynes per cell. Marshall(9) claimed that these values should have been even higher than what was measured.

Zobell(3) was the first to suggest that the firm attachment of bacteria to submerged surfaces was mediated by some sort of "cement" secreted by bacteria. Corpe and co-workers(20-25) observed that marine bacteria which naturally attach to surfaces have the ability to produce extracellular polysaccharides. The chemical composition of these substances seems to be unique

TABLE 1

Variables Affecting the Sorption of Microorganisms by Solid Surfaces(15)

1. Character of the microorganism
   - A. Species
   - B. Culture medium
   - C. Culture age
   - D. Suspension medium
   - E. Concentration
2. Character of the Adsorbent
   - A. Type
   - B. Ionic form (ion exchange resin)
   - C. Particle size
   - D. Crosslinkage (ion exchange resin)
   - E. Concentration
3. Character of the Environment
   - A. Hydrogen ion concentration
   - B. Salt concentration
   - C. Agitation
   - D. Time of contact
   - E. Temperature

for each species, and is thought to at least partially determine the location or habitat that the bacteria occupy in the natural environment. After recovering and purifying these polysaccharides, Corpe found them to be acidic mucopolysaccharides composed mostly of polyanionic carbohydrates. Then he smeared some of the recovered polymer on glass slides and allowed them to air dry. After 3 days he submerged them along with clean slides into both natural and artificial seawater. In both cases the presence of polymer significantly enhanced attachment.

Tosteson and Corpe(25) repeated the same experiment for algal cells. The polymer of the marine pseudomonad caused algal attachment. If the polymer was added to a suspended culture of algae, it caused the cells to clump or aggregate. The same polymer was found by Sieburth(26) to agglutinate arthrobacter cells; this polymer had a very large molecular weight, possibly 5000, but at least 2000.

Marshall et al.(7) observed soft blebs or droplets on the cell surface of attaching bacteria. These were thought to be points of attachment, and it was proposed that attachment is accomplished by extracellular polymeric fibrils which concentrate at the points of attachment. Corpe et al.(24) suggested that attachment may be controlled by material excreted at the cell wall rather than by the abundant mucopolysaccharide which forms the slime. To test this theory they isolated mutants that were unable to produce or could produce only limited amounts of the mucopolysaccharide. The mutants initially attached more slowly, but after 48 hours they had attached to the same or better extent than the original cells.

Jones et al.(27) in an electron microscopic study of bacterial slimes, observed what they described as an extracellular material that formed strands and net-like forms in the slime matrix. The strands were positioned between 1 and 4 μm apart. After a certain period of time (9 days), polymer production appeared to stop and there appeared a void area surrounding each cell in the matrix. They postulated that the void is a sort of microenvironment that acts as a buffer or protection zone. The polysaccharide strands may act as gradients through which diffusible and readily useable nutrients pass.

Fletcher and Floodgate(28,29) attributed attachment to two polymers closely associated with the cell wall; both contained some acidic polysaccharide. They surmised that the firm attachment process occurs in two steps: initial attachment, carried out by a primary polysaccharide, and bond strengthening, as the result of excretion of a secondary polymer after initial firm attachment is made. The primary polysaccharide was found in populations of both naturally attaching bacteria and suspended bacteria. When non-attaching bacteria were forced into contact with a surface by filtration, the primary polymer stretched to the surface just as it did for the naturally attaching bacteria. When naturally attaching bacteria were grown in high pH medium, the production of the primary polysaccharide was almost completely eliminated. This, however, did not interfere with adhesion.

The primary polysaccharide can be broken into two regions according to Fletcher and Floodgate(28,29). An inner dense line on the cell wall surface and an outer fine grained "fringe" region about 15 to 25 nm thick. The secondary polysaccharide is usually only found where there are groups or colonies of bacteria. It is a fibrous substance that stretches around adjacent bacteria or between the bacteria and the surface. They believe that the secondary polysaccharide evolves from the primary. One possible explanation would be the further polymerization of the primary polymer by a substance secreted by the bacteria. A change in pH for instance, could cause such polymerization. Bacteria grown at 25°C exhibited little propensity for attachment. The reason for this is unexplainable based on current knowledge. Fletcher(28,29) suggested one factor may be the weakening of the polymers at the increased temperatures.

When Fletcher and Floodgate developed bacterial films in glucose-artificial seawater, the secondary polysaccharide failed to appear (the primary was retained) and the bacteria rested in "pockets" which appear to be identical to the "buffer zone" of Jones et al.(27). Bacteria grown in peptone, yeast extract and aged seawater formed the secondary polysaccharide but formed no pockets.

Marshall observed that addition of 7 mg/ℓ of glucose to his sewater medium stimulated attachment markedly. The addition of higher amounts, however, almost completely inhibited attachment from occurring, although it stimulated the growth rate. Concentration of glucose over 30 mg/ℓ completely inhibited bacterial attachment. Jones et al.(27) used nutrient broth in river water and after a period of 9 days observed both the secondary polysaccharide and the buffer zone. Perhaps Fletcher terminated the experiments involving the peptone-yeast extract before the formation of the pockets occurred.

When bacterial films were transferred to Ca^{2+} and Mg^{2+} deficient medium, a severe and almost immediate disruption of the secondary polysaccharide occurred. Ca^{2+} and Mg^{2+} are obviously important in the maintenance of the structure of the intercellular matrix. Presumably, the acidic polysaccharides are held together by divalent bonding(29).

Corroborating Fletcher and Floodgate's(2) finding, Marshall(9) observed that if Ca^{2+} and Mg^{2+} were completely eliminated from the medium, irreversible attachment failed to occur. If only one of the divalent ions was omitted from the feed, attachment occurred at the same rate as if both cations were present.

Several researchers have attempted to find a method or treatment to remove bacterial slimes from surfaces. Marshall found that freshwater slimes were sensitive to periodate cleavage. Bacterial films grown in seawater were insensitive to the periodate treatment unless it was preceded by the addition of EDTA. He proposed that the divalent ions in the seawater protected the bacteria from the periodate and that the EDTA acted as a chelating agent of the ions allowing the periodate to become effective. Corpe(23) found that there was a significant loss of attachment by a 5 minute exposure to NaOH, anionic and nonionic detergents, chelating and oxidizing agent and protein denaturing agents.

Phase 2 attachment as can be seen is a very selective process. Marshall (9) observed that bacteria attaching to surfaces usually have rough or convoluted surfaces and dense polar bodies. Corpe(22) reported that of all the marine colonizing bacteria that he isolated, virtually all were small rod shaped and gram negative. From 60 to 90 percent of the isolates were pseudomonads, while the remaining 10 to 30 percent were either Flavobacterium or Achrobacter.

Other factors that effect bioadhesion are the composition of the medium and the characteristics of the surface or substratum. Baier(30,31) has also demonstrated that adsorbed coatings on the submerged surfaces can either increase or decrease attachment. Other investigators(32,33,34) have also studied the relationship between the properties of the solid surface and the permanent attachment of a biological film. The following section explores such a relationship in more detail.

EFFECT OF THE PROPERTIES OF THE SOLID SURFACE ON BIOFILM ATTACHMENT

Since the late 1940's, researchers have attempted to establish a relationship between surface (substratum) properties and biofilm attachment. Weiss(16,35) was the first to introduce the concepts of wettability, contact angle, and critical surface tension to the field of bioadhesion.

Weiss and Blumenson(37) observed that cells attach more readily to high energy surfaces such as glass than they do to materials like teflon, that have low surface energy. Baier(31) was able to reinterpret the results of another researcher in terms of critical surface tension. He examined the percent of cells that "spread" (thought to be a measure of cellular adhesiveness) versus critical surface tension and found an excellent correlation.

Weiss and Blumenson(36) found that if horse serum was added to the medium they used, there was effectively no difference between the attachment of cells to glass and teflon. They concluded that an adsorbed film of serum on the glass and teflon was serving as the "surface" for both materials and therefore attachment should be the same. Other researchers have investigated the effect of various proteins on bacterial attachment, and found that in general, they tend to retard the process. Meadows(13) observed that salmine and albumin inhibited attachment, but casin and gelatin appeared to enhance attachment. Fletcher(32) concluded that proteins could affect attachment in one of three possible ways. The protein could be inhibitory to the bacteria, affect the surface properties of the bacteria or change the surface properties of the substratum. Fletcher(32,33) tested adsorption under varying conditions and concluded that the surface properties of the substratum were being altered. Bovine serum, albumin, gelatin, fibrinogen and pepsin all impaired the attachment of a marine pseudomonad to polystyrene. Basic proteins, protamine and histone had no effect, presumably because they failed to form an adsorbed film at the interphase.

Dexter(34) concluded that in natural systems such as seawater, a submerged surface becomes almost immediately coated with an adsorbed glycoprotein film. He submerged various surfaces in seawater and found a good correlation between the rate of attachment of bacteria and the critical surface tension of the clean substratum. He concluded that although it is unlikely that the surface properties of the pure clean substratum will directly affect the attachment process, they seem to indirectly affect it, possibly by determining the composition of the adsorbed layer or conditioning film. The formation of such an adsorbed film at a platinum surface submerged in seawater was demonstrated by Loeb and Neihof(44). Baier(30,31) suggested that adsorption of an organic monolayer on the substratum conditions its surface, thus becoming a prerequisite for bacterial attachment.

Once it becomes apparent that attachment is controlled to a fairly large extent by surface properties of the substratum, there are two possible alternatives in attempting to control bioadhesion. The substratum surface properties may be altered via substitution of low energy groups, such as fluorine, or a coat of surface modifer could be applied to the substratum. Eirich(41), in talking about the use of polymers as a coating, made some very good points. Because of the many adsorptive sites, polymers are virtually insensitive to variations on localized surface conditions. Furthermore, due to their intrinsic nature, polymers may extend adsorption sites out varying distances from the substratum, building a sizable and interwoven interphase.

In the water and wastewater treatment, surface pretreatment by an artificial polymer coating seems to be a workable approach to enhance attachment of colloids and bacteria. In the water treatment area, several researchers (39,40,41) have demonstrated the effectiveness of polymer precoating in multimedia filters for direct filtration of colloidal suspensions. In the wastewater treatment field, however, very little research has been done on enhancing biofilm growth.

Biofilm reactors, such as the rotating biological contactors, trickling filters and fluidized beds have little or no problem in developing a stable biological film for aerobic BOD removal or for denitrification. However, certain processes, namely, biological nitrification and anaerobic treatment, may require several weeks(42), and even months(43), to develop a healthy growth. It is clear that in these cases it would be desirable to develop a method to enhance biofilm growth.

The information presented in the following sections summarizes preliminary results of the research being carried out at the University of Massachusetts. This research has the overall objective of developing a good understanding of the factors controlling film attachment in wastewater treatment systems. The experiments reported herein were designed to test the effect of substratum precoating with artificial polymers on biofilm growth and substrate uptake rates.

PRELIMINARY EXPERIMENTS

Materials and Methods

This preliminary experimental phase was conducted at the University of Massachusetts Wastewater Treatment Pilot Facility, which is located adjacent to the Town of Amherst Wastewater Treatment Plant. Films of heterotrophic microorganisms were grown in the first four runs, using primary effluent from the latter. TOC in this wastewater ranged from 16 to 65 mg/ℓ, TSS averaged 55 mg/ℓ, and NH_4^+-N averaged 13 mg/ℓ. The last experiment was run using secondary effluent from an extended aeration pilot unit, to which supplemental ammonium, alkalinity and hardness were added to enhance growth of a biofilm of nitrifying microorganisms.

The experiments were done using a twin-chamber rotating disk reactor. This plexiglass bench-scale unit was operated under complete-mix continuous flow conditions. Its two chambers work in parallel, one as the experimental unit, and the other as a control. Each compartment is 7.1 cm wide, 17.3 cm long, and 19 cm deep. A 14-cm diameter plexiglass disk (1-cm thick) rotates with its axis perpendicular to the 17.3 cm x 19 cm chamber wall. The rotational speed, maintained at 120 rpm, provided for complete-mix conditions, and was deemed appropriate to eliminate external mass transfer resistances. The water depth was maintained at 17 cm to keep the disk completely immersed.

In order to sample the biofilm growing adhered to both the reactor walls and the disk surface, each chamber is provided with eight removable slides, four on the walls, and two on each side of the disk. Film thickness determination was made optically, using a microscope equipped with a stage micrometer. First, a narrow portion of the biofilm was scraped away and a pencil mark was made on the slide surface. Then, a small piece of cover glass, also marked with a pencil, was placed on top of the biofilm, over the mark on the slide. Finally, the film thickness was determined by first focusing the microscope on the mark on the slide, and then on the mark on the piece of cover glass; the difference of the two readings gave the film thickness in microns. This procedure was repeated several times in one slide from the disk and one slide from the wall. An area-weighted average film thickness was computed and plotted versus time in the first four runs.

Since there are quite a number of different polymers and concentrations that could be used to precoat the surfaces of the biological reactor, available polymer samples were tested systematically at different concentrations to determine which are most suitable to enhance biological film growth. The polymer samples used in this study were commercially available polymers, obtained from several companies which make chemicals specifically for water and waste water treatment applications (e.g. Nalco 610, Nalco 7120, Calgon WT 2640, Senfloc 140, American Cynamid 2535, Percol 722).

For each polymer tested, an optimum concentration of polymer was first determined by dissolving an arbitrary amount of polymer in distilled water; from this solution a series of six dilutions was then prepared. Next, six pairs of microscope slides were soaked in each
24 hours. After being air dried, the slides were placed into the six 1-liter beakers of a jar test apparatus, each of which was filled with secondary effluent. The growth of biological film on each slide was then visually monitored for up to 48 hours to see what effect, if any, the polymer precoating had. If there was little or no growth, another test was tried to see if a higher polymer concentration would be more effective.

The optimum polymer concentration, i.e., that which gave the largest visible amount of growth, was then used to coat the experimental chamber. This compartment was filled with the polymer solution, left overnight, drained the next day, and then allowed to air dry; the other chamber was filled with distilled water.

After the exposed surfaces in the experimental chamber were dry, each experiment was started by running primary sewage or secondary effluent continuously, at the same flow rate, through both parallel chambers. The flow rate in the first run was 10 ml/min, in the next three runs, 12.5 ml/run, and in the last run, 35 ml/min.

The system was kept aerobic by bubbling air into the chambers. Dissolved oxygen levels in the first four runs were kept above 3.0 mg/ℓ. The DO level in the fourth run, however, ranged between 1.5 and 2.5 mg/ℓ, while in the fifth run it remained under saturation conditions. Other environmental factors, such as pH and temperature were maintained within acceptable limits; pH ranged from 6.0 to 7.0 in all runs, while the temperature in each individual run were the following: 20^{o}-22^{o}C in Run 1, 19^{o}-20^{o} in Run 2, 18^{o}-19^{o}C in Run 3, 17^{o}C in Run 4, and 11.5^{o} to 16^{o}C in Run 5.

Variables monitored in each of the first four runs include film thickness and TOC removed. In the last run, the concentrations of ammonium and nitrate, as well as film thickness were continuously monitored. The results are presented in the following section.

Results and Discussion

The first three experiments were run using a polymer precoat on all the exposed surfaces of one of the chambers. The polymers used were American Cynamide 2535, Percol 722; and Calgon WT 2640 for runs 1, 2 and 3 respectively; all are cationic polymers, proven to be effective in wastewater sludge conditioning.

Results of these first three experiments are presented in Figures 1 through 3. Figure 1 shows that American Cynamic 2535 had a significant effect in accelerating biofilm growth until the fifth day. After the sixth day the chamber without polymer precoat showed a thicker biofilm throughout the exposed surfaces. At the seventh day the biofilm in the precoated chamber began sloughing off, while no significant film sloughing could be noticed in the control compartment.

Physical appearance of both films was also different. While the control chamber exhibited a fluffy, slimy film, the chamber with precoat had a more compact and uniform film. This difference could be due to simultaneous coagulation of inorganic colloids and bacteria on the polymer layer. However, no significant difference on living biota could be observed under the microscope between the two biofilms.

The same distinctions in physical appearances were observed when other polymers were used. Figure 2 shows the results obtained with Percol 722; in this case there was a marked improvement in biofilm growth. On the first day, the film in the chamber with precoat was 3.2 times thicker than that in the control unit, while on the fifth day, it was 2.5 times thicker. Sloughing of the thicker film begun at the fifth day, and continued until the end of the run, while the control unit showed no significant sloughing. The reason for this could be that the thinner film had not attained a sufficiently large thickness for sloughing to occur.

Since the largest observed difference in film thickness in runs 1 and 2 took place at the first day, in run 3 more frequent observations were made during the first 24 hours. Figure 3 shows that after 5 hours, measurable growth had still not occurred on the uncoated surfaces, while there was a measurable biofilm thickness on the coated surfaces. Biofilm growth was still unmeasurable after 18 hours in the control chamber; at 23 hours, only 22 microns had accumulated, while the coated chamber had 6.3 times that amount. From the beginning until the third day, the chamber with precoat had a thicker film than the control one. By the fourth day there was no significant difference between both film thicknesses.

The results of experiments 1 through 3 show that no polymer precoat is needed to grow a biofilm in an aerobic wastewater treatment unit which removes organic matter from primary effluent. These runs, however, demonstrate that a significant improvement on biofilm development could possibly be achieved by precoating the exposed surfaces in treatment processes in which it is difficult to grow a biological film, nitrification and anaerobic biofilms being the most notorious examples. Thus, the final experiment was designed to test the effect of coating the exposed surfaces of one of the reactor chambers with Calgon WT 2640 (the most efficient of the polymers tested in runs 1-3), on the development of a film of nitrifiers.

Preceding this experiment, another run was necessary. Since the Amherst wastewater is low in total hardness, in view of Fletcher and Floodgate's (29) and Marshall's (9) observations, it was necessary to test if adding supplemental hardness would have any effect on biofilm development. Thus in experiment 4 calcium chloride was added in such amounts that the total hardness, which usually averages 80 mg/ℓ as $CaCO_3$ or less, would be increased to more than 300 mg/ℓ as $CaCO_3$. No polymer precoat was applied to either chamber; one of them received primary effluent with extra calcium while the other served as a control unit.

Figure 4 shows the results of this run. No growth could be measured for the first two days; however, on the third day the chamber with extra calcium had developed a film more than twice as thick as that in the control unit. On the fourth day the ratio of film thicknesses decreased to 1.3, and by the fifth day, film thicknesses in both chambers were virtually equal. It was clear, therefore, that adding supplemental calcium had a positive effect on early biofilm development. For this reason, it was decided to correct the calcium deficiency of the wastewater for the last experiment.

The fifth run was performed using final effluent from an extended aeration activated sludge pilot unit. Ammonium bicarbonate was added to bring the influent ammonium concentration to more than 25 mg/ℓ as N. Total hardness was increased to about 300 mg/ℓ as $CaCO_3$, and sodium bicarbonate was added to provide sufficient alkalinity for the nitrification reaction.

By the end of the first day of the final run the walls of the precoated chamber had developed a biofilm which did not completely cover the exposed surfaces. While roughly 10 percent of the wall area showed attached growth, the rest of the walls, and the rotating disk did not have any measurable biofilm. Nevertheless, this apparently small film development was able to carry out roughly 12 percent of ammonium removal. On the other hand the uncoated chamber, which had no measurable growth, showed no removal whatsoever.

The percentage of the wall surface covered with biofilm kept increasing until the end of the run. By the end of the sixth day, approximately 50 percent of the wall surface had been covered. The disk surfaces, on the other hand, did not develop any measurable growth. This was probably due to the higher shear forces affecting such surfaces.

The plot presented on Figure 5 shows the erratic biofilm growth observed on the removable slides of the chamber walls. Significant sloughing could be noticed, as the amount of suspended solids kept increasing with time. As indicated in the same plot, the uncoated chamber did not exhibit any measurable biofilm development.

Figure 6 shows the percentage of ammonium removal in both chambers. It can be seen that whereas the uncoated reactor did not remove any ammonium from solution, the coated chamber showed a clear improvement in ammonium removal. However, the percentage removed was inconsistent with the observed film development. The reason for this behavior could be that the overall ammonium removal was due to both the attached growth and the suspended growth (sloughings). In fact, the higher removals coincide with the largest decrease in film thickness, i.e., between the first and the third days, in which the suspended solids increased. It was observed that the sloughings remained in suspension for a long time, thus contributing to the observed ammonium removal.

The last experiment demonstrated that precoating the exposed surfaces with Calgon WT 2640 was ineffective in promoting complete film development. Nevertheless, considering that the water temperature was considerably lower than the optimum for nitrification, the partial success achieved in this run demonstrates that it is feasible to accelerate the development of a film of nitrifiers by precoating the exposed surfaces with an appropriate polymer. Finding such a polymer is under investigation at the University of Massachusetts.

SUMMARY AND CONCLUSIONS

Although sanitary engineers have used biofilm reactors for more than 100 years, very little research has been done to develop a good understanding of the factors affecting biofilm development. Most of the research performed in this area has been conducted by marine microbiologists with the aim of preventing fouling of surfaces exposed to seawater.

The concensus is that besides factors such as surface to volume ratio and substrate concentration, there are many others which affect bioadhesion. It has also been generally accepted that the development of a stable biofilm occurs in two phases: reversible adsorption and permanent attachment.

The first phase, which takes place within a short lapse, does not yield permanent cell attachment. In fact, at this stage, the cells can be removed by mild rinsing.

The second phase, on the other hand, results in firm cell attachment. To achieve a permanent bond, the cells excrete polysaccharides of different kinds. These polymers, when extracted and applied to either solid surfaces or to suspended cells, enhance biofilm development or promote coagulation of the suspended growth.

The surface properties of the substratum appear to have a very important role on the rate biofilm development, surfaces with low energy taking longer in allowing cell adhesion. However, it must be emphasized that in many instances adsorption of a layer of solute on the substratum can radically change its surface properties.

Based on this premise preliminary experiments were designed to test the effectiveness of precoating the exposed surfaces with commercially available polymers on the development of a biofilm of nitrifiers, which is usually difficult to grow.

The first three experiments were designed to select the most effective polymer. The fourth run was carried out to determine the effect of total hardness on film development. Finally, the fifth run was intended to demonstrate the effect, if any, of precoating the exposed substratum surfaces. The following conclusions can be drawn from results obtained in these five preliminary experiments:

1. The bench scale unit used in this research proved to be adequate for selecting the best polymer for precoating the exposed surfaces of the substratum.

2. Precoating the exposed surfaces with synthetic cationic polymers can significantly enhance initial film attachment.

3. Film growth in low-hardness waters can be enhanced by addition of divalent cations.

4. Whereas no nitrification was observed during six days in the uncoated chamber, a low level of nitrification could be observed, as early as within 1 day, in the chamber precoated with Calgon WT 2640.

5. Results of these preliminary experiments demonstrate that it is feasible to accelerate the development of a film of nitrifiers by precoating the exposed substratum surfaces with a suitable polymer. A more extensive search among the commercially available polymers may prove fruitful.

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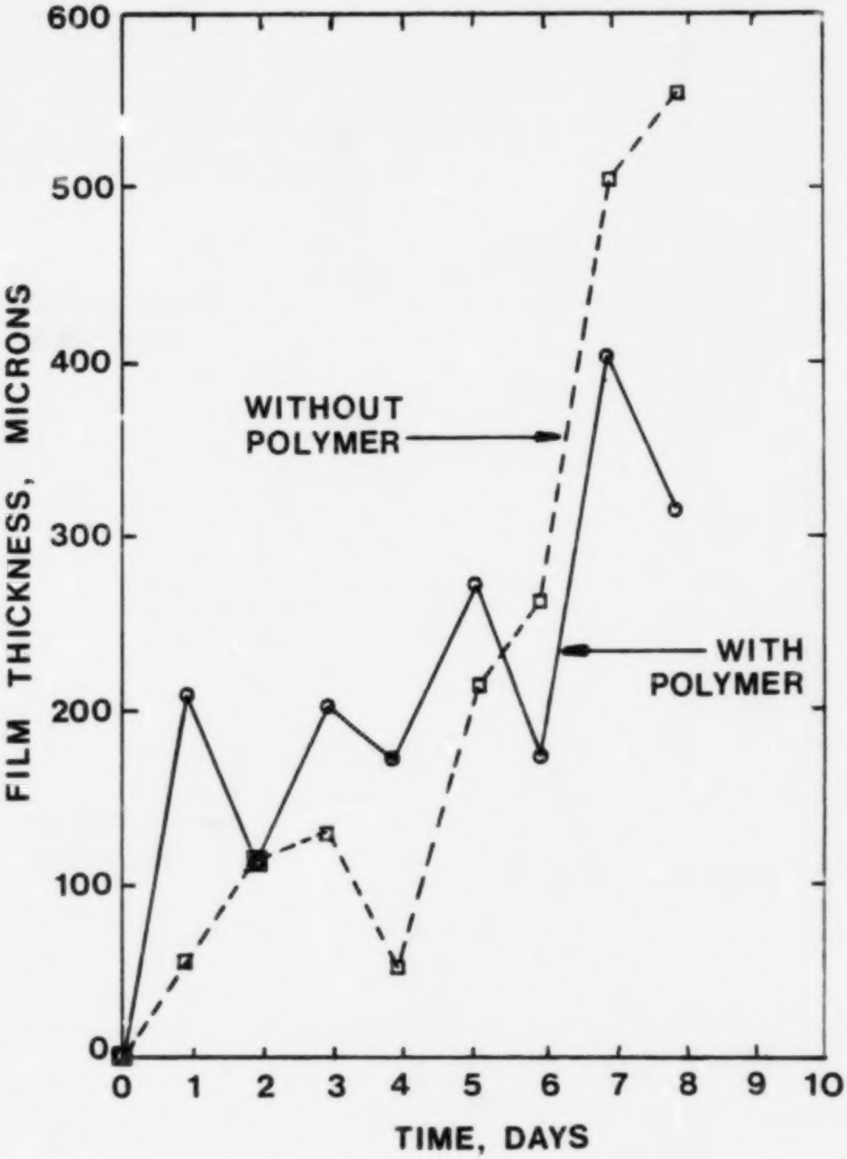

Figure 1. Comparison of film growth in chamber precoated with American Cynamid Polymer 2535 and in chamber without precoat.

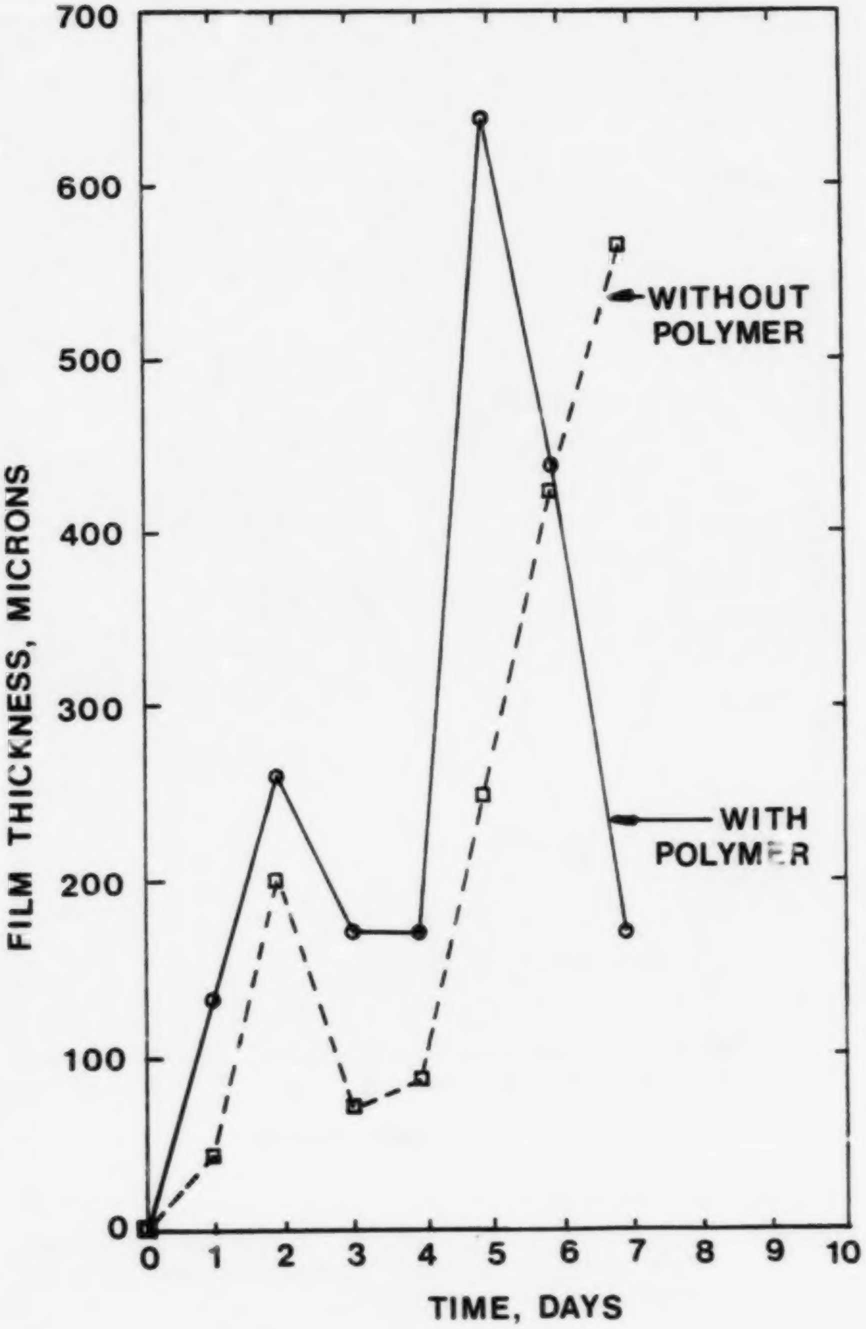

Figure 2. Comparison of film growth in chamber precoated with Percol 722 and in uncoated chamber.

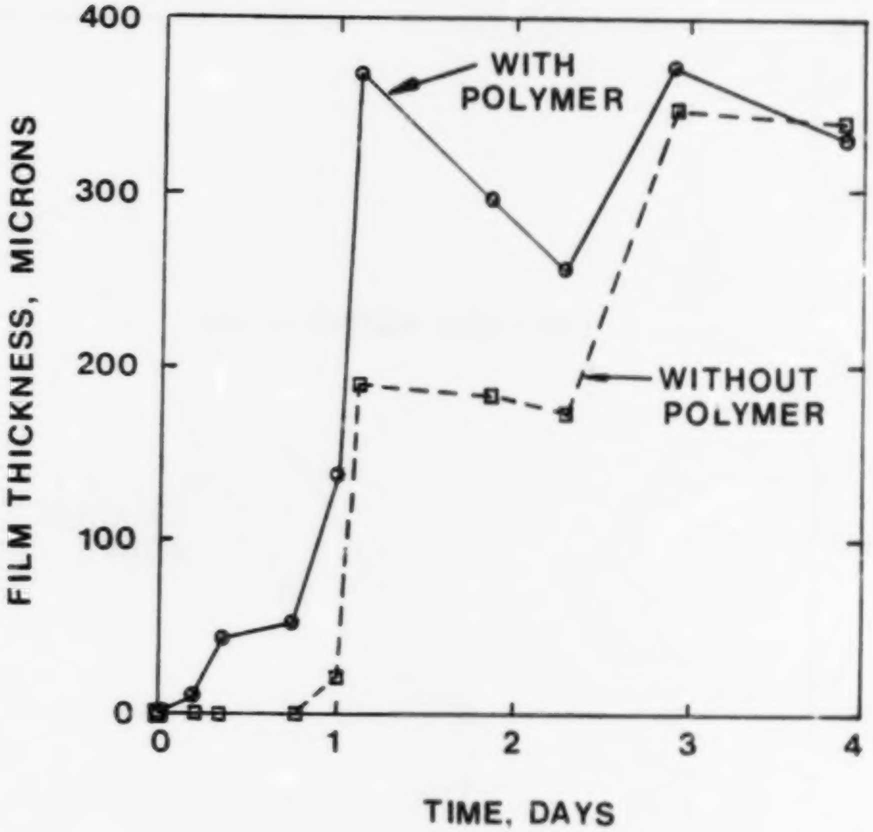

Figure 3. Comparison of film growth in chamber precoated with Calgon WT 2640 and in uncoated chamber.

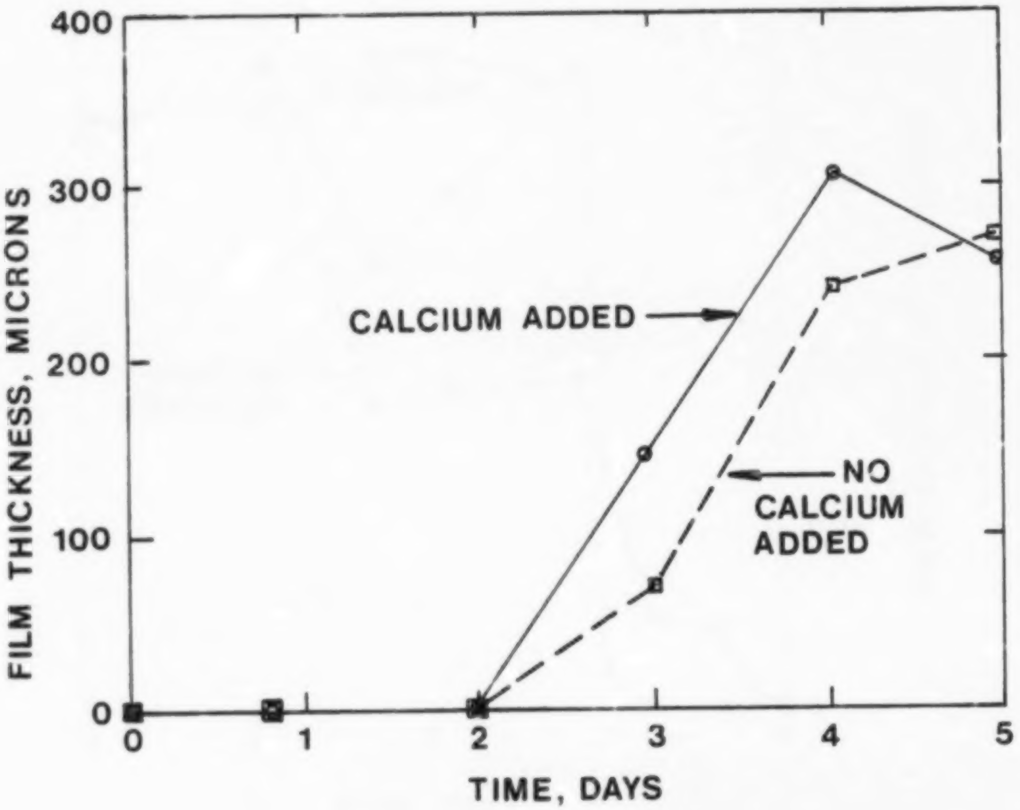

Figure 4. Effect of adding supplemental hardness on film growth. No polymer precoat was applied to either chamber.

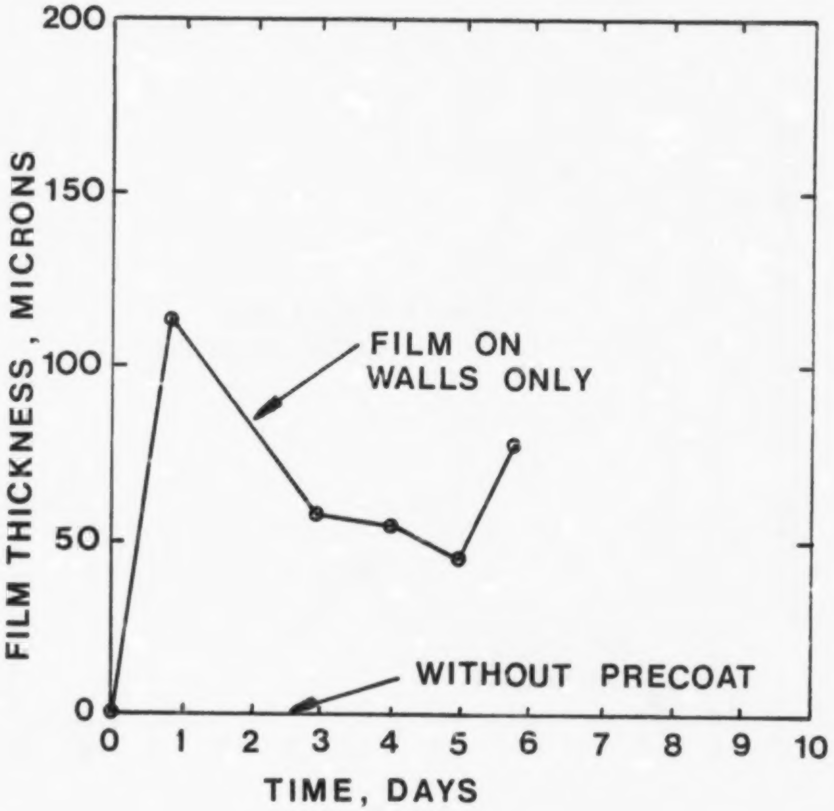

Figure 5. Growth of film of nitrifiers in chambers with a precoat of Calgon WT 2640 and without precoat.

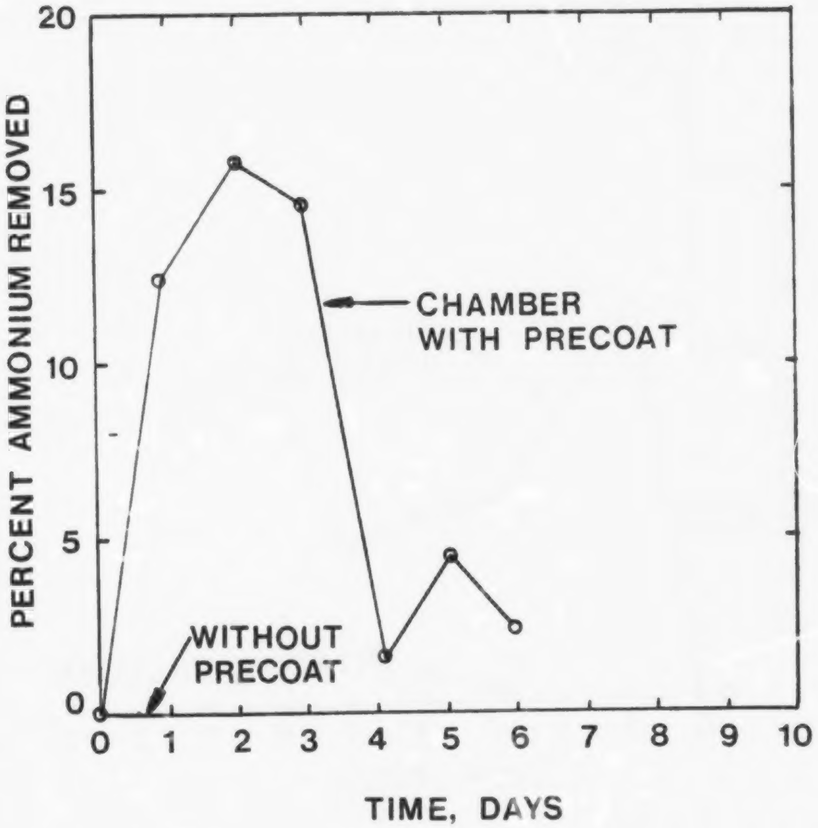

Figure 6. Ammonium removal in chambers with a precoat of Calgon WT 2640 and without precoat.

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A CONCEPTUAL MODEL OF RBC PERFORMANCE

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Introduction

An important task of an environmental engineering educator is to delineate the fundamental principles underlying the operations and processes utilized in our field. The reason for this is two-fold. First, as teachers, by focusing on fundamental concepts we can give our students a depth of understanding which will allow them to analyze new problems which go beyond the illustrations and examples covered in the classroom. Second, as researchers, we can make our work more broadly applicable than it would be if we focused only on specific applications of a process. One tool which can augment our focus on fundamental concepts is mathematical modeling. Again, the reason for this is two-fold and is related to the dual role of the engineering educator. First, and most importantly, the act of developing a mathematical model forces the developer to attempt to describe all of the phenomena occurring within the process, thereby making him aware of the limitations of knowledge and requiring him to deal with them, both by expanding the knowledge base through research and by making all assumptions explicit so that the limitation of the model are evident. Second, fundamental mathematical models provide a mechanism for extending our knowledge into areas that haven't been covered in practice.

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All of the reasons outlined in the preceding paragraph are particularly important to an analysis of RBC's. Again, there are two reasons for this. First, our experience with them is limited so that it is difficult to extract "universal truths" about their behavior from the literature available. Second, they are extremely complex, so that the extrapolation of empirical data to other situations is very risky, at best. Consequently, as teachers, we decided that our students would benefit most from an analysis of RBC's based upon fundamental concepts and therefore we developed the model presented herein(1). Since we were not concerned with immediate application, but rather with education, our guiding principle was the illustration of the various parallel events occurring within these systems. Even then, we came upon limitations in our knowledge and abilities which neccessitated compromises and simplifying assumptions. Nevertheless, in the spirit of this symposium we want to share our thoughts in the hope that some of them may be useful to those involved in actual RBC experimentation and application.

MASS TRANSFER RESISTANCES AND THEIR EFFECTS

The major characteristic of an RBC, that the microorganisms grow attached to a solid surface which is moved through the liquid undergoing treatment, is the very thing which makes its analysis and description difficult. This follows from the fact that rates of microbial reactions are determined by the concentration of substrate surrounding the microbes. In a homogeneous reactor like a continuous stirred tank reactor (CSTR) the microbial cells are uniformly dispersed so that the substrate concentration to which they respond is the same as the concentration in the liquid. In a heterogeneous reactor like a RBC, on the other hand, the requirement for movement of the organic matter from the liquid phase into the biofilm causes the substrate concentration surrounding the microorganisms to be less than the concentration in the bulk liquid. This means that it is necessary to combine physical mass transport with microbial reactions when modeling an RBC. When this is done, two resistances to mass transport must be considered, one in the liquid film adjacent to the biofilm, and the other within the biofilm itself. In order to illustrate the importance of each of these and the technique for handling each, we will first consider them individually. After that, their combined effects will be considered.

External Mass Transfer Resistance

Consider the ideal case of a flat plate covered with a mono-layer of microorganisms so that substrate is consumed entirely at the liquid-biofilm interface. Because of this consumption the concentration of substrate at that interface will be smaller than the concentration in the bulk liquid, as shown in Figure 1a. In order for consumption to continue, substrate must be transported from the bulk fluid to the interface by diffusive and convective mass transfer processes. Consequently, the observed reaction rate depends upon the rate of mass transport to the biofilm as well as on the true, intrinsic rate of reaction at the biofilm. Only when the mass transfer resistance is negligible will the observed reaction rate be equal to the intrinsic kinetics of the reaction; for all other situations the external mass transfer resistance will obscure the intrinsic reaction rate.

In order to model this situation we may idealize the substrate concentration profile by confining it to a hypothetical stagnant liquid layer as shown in Figure 1b. The rate of mass transfer of substrate from the bulk fluid across a unit area of the liquid boundary layer to the biofilm is called the flux, N_s, and is proportional to the change in substrate concentration across the stagnant layer:

$$N_s = k_L\ (S_b - S_s) \tag{1}$$

The proportionality constant, k_L called the mass transfer coefficient, incorporates all of the effects of diffusive and convective mass transport into one parameter. Thus, its value depends upon the properties of the fluid, the diffusivity of the substrate in the fluid, and the nature of the turbulence, which can be represented in part by the bulk fluid velocity. Because of the importance of mass transfer to many engineering problems considerable effort has gone into the development of equations which allow the prediction of k_L under a variety of conditions. A detailed treatment of the results of these efforts may be found elsewhere (2).

The consumption of substrate by the biofilm may be assumed to follow Monod (3) kinetics:

$$-r_s = \frac{q_m\ S_s}{K_s + S_s} \tag{2}$$

In this situation, where the cells are attached to a solid surface, it is advantageous to express the reaction rate in terms of a unit surface area rather than a unit mass of cells. In this way it is assumed that each unit area of biofilm contains the same unit mass of active cells. Thus $-r_s$ is the mass of substrate consumed per unit time per unit surface area of biofilm and q_m is the maximum possible removal rate per unit area (which incorporates the cell mass per unit area). K_s is a saturation constant so that the term $S_s/(K_s + S_s)$ causes the removal rate to decline in a hyperbolic fashion as the substrate concentration declines.

The problem with a rate equation like Eq. 2 is that S_s, the concentration of substrate at the surface of the biofilm, is difficult to measure. Thus it would be advantageous to express the rate in terms of the bulk concentration, S_b which is easy to measure. When a steady state exists at the surface of the biofilm there is no accumulation of substrate and hence the rate of substrate supply through the stagnant liquid layer by mass transfer must equal the rate of substrate consumption, i.e. N_s must equal $-r_s$. Thus,

$$k_L\ (S_b - S_s) = q_m\ S_s/(K_s + S_s) \tag{3}$$

Equation 3 may now be solved for S_s in terms of S_b and the result substituted into Eq. 2 to obtain an expression for the reaction rate as a function of the bulk substrate concentration:

$$-r_s = \frac{q_m\ \{S_b - K_s - q_m/k_L + [(S_b - K_s - q_m/k_L)^2 + 4K_sS_b]^{0.5}\}}{S_b + K_s - q_m/k_L + [(S_b - K_s - q_m/k_L)^2 + 4K_sS_b]^{0.5}} \tag{4}$$

It is clear from Eq. 4 that the presence of external mass transfer resistance prevents the rate expression from exhibiting a Monod-type dependence on the bulk substrate concentration, even when the true kinetics do so.

Although Eq. 4 demonstrates that the existance of an external mass transfer resistance changes the form of the equation for the substrate removal rate, the exact effect of that resistance is difficult to visualize. In chemical engineering this problem has been reduced through use of the effectiveness factor, η, which is defined as the ratio of the actual, observed reaction rate to the theoretical rate that would occur in the absence of mass transfer resistance. In this way the actual reaction rate can be obtained by multiplying the theoretical rate at the bulk substrate concentration by η. For Monod kinetics at the biofilm, the substrate removal rate would be given by

$$-r_s = [q_m S_b/(K_s + S_b)]\, \eta_e \tag{5}$$

where the subscript e on η_e reflects the fact that the effectiveness factor is correcting for external mass transfer resistance.

The equation for the effectiveness factor as a function of the system parameters may be obtained by equating Eq. 4 with Eq. 5 and solving for η_e. The result shows that η_e depends upon the bulk substrate concentration, S_b, as well as upon the kinetic parameters q_m and K_s and the mass transfer coefficient, k_L. Thus, in order to make the expression for the effectiveness factor of general utility, dimensionless variables are used:

$$\bar{K} = K_s/S_b \tag{6}$$

$$D_a = q_m/k_L\, S_b \tag{7}$$

The dimensionless group, D_a, is called the Damköhler number and it is the ratio of the maximum possible substrate removal rate at the interface (q_m) to the maximum possible substrate transfer rate across the stagnant liquid layer ($k_L S_b$). When $D_a > 1$, the maximum substrate removal rate exceeds the maximum rate of substrate transport and the situation is said to be transport-limited. When $D_a < 1$, the opposite is true and the situation is said to be reaction-limited. Using these groups, the equation for η_e can be shown to be:

$$\eta_e = \frac{(\bar{K} + 1)\,\{1 - D_a + [(1 - \bar{K} - D_a)^2 + 4\bar{K}]^{0.5}\}}{(1 + \bar{K} - D_a) + [(1 - \bar{K} - D_a)^2 + 4\bar{K}]^{0.5}} \tag{8}$$

The effectiveness factor is plotted against D_a in Figure 2 using $\bar{K}$ as a parameter. When D_a is very small the effect of mass transfer is negligible and the effectiveness factor is approximately equal to unity. In other words, when the maximum rate of mass transport is much greater than the maximum reaction rate, the rate of substrate removal can be expressed by the Monod equation with substrate concentration equal to that in the bulk of the fluid because the substrate concentration at the biofilm will be approximately equal to that concentration. As D_a exceeds unity and the maximum mass transfer rate becomes slower with respect to the maximum substrate removal rate the effectiveness factor decreases sharply, thereby demonstrating that mass transfer has become quite important in determining the performance of the system. Under those circumstances the rate of substrate removal must be determined with Eq. 5 because the substrate concentration at the biofilm is less than the concentration in the bulk of the fluid. Furthermore, Figure 2 shows that a decrease in $\bar{K}$ also acts to decrease η_e. This follows from the facts that a decrease in $\bar{K}$ makes K_s smaller with respect to S_b and that a decrease in K_s acts to increase the rate of substrate removal by the biofilm.

For a given set of biokinetic parameters the net result of a mass transport limitation is to increase the bulk substrate concentration with respect to that which would exist in the absence of the mass transport limitation. Furthermore, because η_e depends upon S_b the magnitude of the effect will change as the bulk substrate concentration is changed. This is illustrated in Figure 3 which shows the calculated effluent substrate concentration from a CSTR containing a mono-layer of bacteria of fixed surface area. The lower curve represents the case of no mass tansfer resistance (i.e. intrinsic kinetics) whereas the upper curve represents moderate resistance. When the liquid flows through the reactor at a rate which produces a substrate concentration of 20 mg/l in the absence of mass transfer resistance, the imposition of such a resistance will increase it to 70 mg/l. On the other hand, when the liquid flow rate is such that a concentration of 2 mg/l would result without mass transfer resistance, its presence would increase S_b to only 5 mg/l. The reason for the differential magnitude of the mass transfer effect can be seen by reexamining the parameters in Figure 2. As S_b is decreased, both D_a and $\bar{K}$ will increase. However, their effects on η_e are opposite so that the increase caused by $\bar{K}$ tends to off-set somewhat the decrease caused by D_a. The magnitude of this difference will depend upon the relative values of the parameters, making generalizations difficult. Nevertheless, the utility of the effectiveness factor concept in illustrating this phenomenon is apparent.

Internal Mass Transfer Resistance

Having seen the effect of mass transfer up to the biofilm and the use of effectiveness factors in expressing that effect, let us now turn our attention to the situation in which substrate must be transported to microorganisms living within a biofilm. In order to focus on what is happening within the biofilm we will assume that there is no external resistance limiting the transport of substrate from the bulk liquid to the liquid-biofilm interface.

Consider a flat solid support covered with a film of microorganisms of thickness L_f as shown in Figure 4. Although the organisms at the liquid-biofilm interface are exposed to substrate at concentration S_b the organisms within the film are exposed to lower concentrations because of the need for substrate to be transported to them by diffusion. The transport process can be characterized by Fick's law, but in this situation the diffusivity must be an effective diffusivity, D_e, which accounts for the retardant effect of the gelatinous matrix on substrate transport:

$$N_s = -D_e \, dS/dx \qquad (9)$$

Techniques for estimating D_e have been described elsewhere (2). At steady state the biofilm will attain a constant thickness and there will be no accumulation of substrate within the film so that a mass balance on substrate within a differential element of film yields:

$$-D_e\ A_s\ \frac{ds}{dx}\bigg|_x + D_e\ A_s\ \frac{ds}{dx}\bigg|_{x+\Delta x} + r_s\ a_s\ A_s\ \Delta x = 0 \tag{10}$$

where A_s is the surface area normal to the direction of diffusion, x is the distance into the film from the support, and a_s is the surface area of microorganisms per unit volume of microbial mass. Substituting Eq. 2 for r_s, dividing both sides by A_s and Δx, and taking the limit as Δx approaches zero yields

$$D_e\ \frac{d^2S}{dx^2} - \frac{a_s\ q_m\ S}{(K_s + S)} = 0 \tag{11}$$

which must be solved with two boundary conditions:

$$S = S_b \qquad \text{at} \qquad x = L_f \tag{12}$$

$$\frac{ds}{dx} = 0 \qquad \text{at} \qquad x = 0 \tag{13}$$

Before discussing the solution of this equation let us express it in terms of dimensionless variables:

$$\bar{S} = S/S_b \tag{14}$$

$$z = x/L_f \tag{15}$$

$$\kappa = S_b/K_s \tag{16}$$

So that

$$\frac{d^2\bar{S}}{dz^2} - \frac{\phi^2\ \bar{S}}{1 + \kappa\bar{S}} = 0 \tag{17}$$

$$\bar{S} = 1 \quad \text{at} \quad z = 1 \tag{18}$$

$$\frac{d\bar{S}}{dx} = 0 \quad \text{at} \quad z = 0 \tag{19}$$

The parameter κ indicates the degree to which Monod kinetics deviate from first order, i.e. a small value of κ indicates practically first order behavior while a large value indicates nearly zero-order behavior. The parameter ϕ is called the Thiele modulus and is defined as

$$\phi = [a_s\ (q_m/K_s)\ L_f^2/D_e]^{0.5} \tag{20}$$

Its physical significance may be seen most easily by squaring and rewriting it

$$\phi^2 = a_s\ (A_s L_f)(q_m/K_s)\ S_b/[(A_s/L_f)\ D_e\ S_b] \tag{21}$$

The numerator represents a maximum first order reaction rate while the denominator represents a maximum diffusion rate. Thus, the Thiele modulus for reaction within a biofilm is analogous to the Damköhler number for reaction at the surface of a biofilm. Like the Damköhler number, a large value of the Thiele modulus represents a situation which is diffusion-limited whereas a small value represents a reaction-limited situation.

Because of the non-linear nature of Monod kinetics, numerical techniques must be employed to solve Eqs. 17-19. Such techniques are tedious and time-consuming. Consequently, for general use we would like to minimize their use. This can be done by using the effectiveness factor concept. Because Eqs. 17-19 are expressed in terms of dimensionless variables, their solution gives a generalized substrate concentration profile through the biofilm. Since the rate of substrate consumption at each point in the biofilm is a function of the substrate concentration at that point, the substrate concentration profile allows calculation of the average reaction rate within the film by integrating over the entire film depth. Furthermore, the effectiveness factor concept allows the average reaction rate to be defined in terms of the bulk substrate concentration and an internal effectiveness factor, η_i:

$$(-r_s)_{avg} = [q_m S_b/(K_s + S_b)]\ \eta_i \qquad (22)$$

Thus by equating the average reaction rate obtained from the numerical solution of the differential equation to the average reaction rate obtained with Eq. 22 it is possible to get a generalized expression for the effectiveness factor as a function of ϕ, κ, and S_b. Such an expression could then be used to account for the effects of mass transport within the biofilm without having to solve the differential equation for each situation. Pitcher (4) has done this and the results are presented in Figure 5 where the internal effectiveness factor is presented as a function of a generalized modulus (as suggested by Bischoff(5)):

$$\phi_p = [\phi\kappa/(1 + \kappa)][2\kappa - 2\ \ln(1 + \kappa)]^{-0.5} \qquad (23)$$

At low values of ϕ_p (<0.5) the system is reaction-limited so that η_i approaches unity, indicating that diffusional resistance is negligible with respect to the maximum reaction rate. When ϕ_p is high (> 3), diffusion strongly influences the reaction rate and the effectiveness factor is inversely proportional to ϕ_p and independent of κ. The major effects caused by Monod-type kinetics are seen at the intermediate values of ϕ_p. There it can be seen that the closer the reaction kinetics approach first order (i.e. the lower the value of κ) the greater the diffusional effects will be.

In order to illustrate the effect of internal mass transfer resistance Figure 6 was prepared in a manner similar to that used to prepare Figure 3. In this case, however, a thick film was used without any external resistance. It should be noted that the lower curve in Figure 6 (i.e. the one without any mass transfer resistance) is identical to the lower curve in Figure 3. A value of D_e of 0.025 cm^2/hr is in the range of glucose in water whereas the smaller value represents retarded diffusion such as might be seen in a biofilm. Just as in the case of the external mass transfer resistance, the magnitude of the mass transport effect depends upon the bulk substrate concentration. In this case, however, the change in the effect as S_b is changed is not as large since only κ is influenced by S_b.

Combined Internal and External Mass Transfer Resistances

Having seen the effects that each of the individual mass transfer resistances can have, we now need to consider their combined effects in order to have an effectiveness factor which will be applicable to the biofilms in RBC's. Thus, consider the situation depicted in Figure 7, in which substrate must be transported across a hypothetical stagnant liquid layer before being carried into the reactive biofilm by diffusion. The approach which must be taken here is analogous to that used with internal resistance alone, i.e. the mass balance equation must be written for a differential element within the biofilm and the resulting differential equation solved to obtain the substrate concentration profile. That profile is then used to determine the average reaction rate which is equated to an average reaction rate as given by Eq. 24

$$(-r_s)_{avg} = [q_m\ S_b/(K_s + S_b)]\ \eta_o \tag{24}$$

in order to determine the functional relationship for the overall effectiveness factor, η_o. Once this has been done, it is no longer necessary to solve the differential equation for each particular situation; Eq. 24 may be used instead.

The mass balance equation on substrate in the differential element of biofilm is the same as Eq. 11 and the boundary condition at x = 0 is the same as Eq. 13. The boundary condition at x = L_f is different, however, because the substrate concentration at the liquid-biofilm interface is not known. Consequently, it must be expressed in terms of the transport of substrate across the stagnant liquid layer:

$$N_s = D_e \frac{ds}{dx} = k_L\ (S_b - S_s) \quad \text{at} \quad x = L_f \tag{25}$$

Fink et al. (6) solved Eqs. 11, 13, and 25 as a special limiting case of more general boundary conditions using the following dimensionless quantities:

$$\psi = k_L\ L_f/D_e \tag{26}$$

$$\phi_f = \phi[1/(1 + \kappa)]^{0.5} \tag{27}$$

where ψ is a Sherwood number and ϕ_f is a modified Thiele modulus. Their results are depicted in Figure 8. Each family of curves is for a given Sherwood number, i.e. for a given set of physical conditions determining the magnitudes of the internal and external mass transfer resistances. The family of curves for $\psi = \infty$ represents the case where external mass transport is very rapid with respect to internal mass transport so that only internal mass transport limits the movement of substrate. Lower values of ψ represent the situations in which external mass transfer resistance plays a greater role. Within each family of curves, the value of the Thiele modulus, ϕ^2, signifies the relative importance of reaction rate to diffusion in limiting the system. The larger the value, the more diffusion limits the removal of substrate. The fact that the families of curves are more widely separated than the curves within a family implies that external mass transfer resistance plays a more important role than internal mass tansfer resistance in limiting the rates of substrate removal by fixed-films. The effect of the bulk substrate concentration, S_b, upon η_o is accounted for through its effects on κ.

Although the information in Figure 8 could be used to generate curves showing the combined effects of internal and external mass transfer resistance on the performance of a CSTR containing a microbial film, the results would be qualitatively similar to those in Figures 3 and 6. Thus such a figure was not included.

MODEL OF AN RBC WITH A SINGLE LIMITING SUBSTRATE

Having established a technique whereby both internal and external mass transfer effects may be accounted for, we are now ready to develop a conceptual model for an RBC. In order to demonstrate the technique and to show as clearly as possible what the effects of various physical factors are, this will be done first for a single rate-limiting substrate. Consequently, the electron donor (i.e. carbon and energy source) will be considered to be rate-limiting and all other nutrients, including the electron acceptor, will be considered to be present in excess. After the concepts have been established it will be a relatively straight-forward matter to extend the model to the situation in which both the electron donor and acceptor may be rate limiting.

Model for Substrate Removal in a Single RBC

Because of the complex nature of an RBC, certain simplifying assumptions must be made to allow it to be modeled with reasonable effort.

1. The liquid in the tank housing the discs is completely mixed so that the concentration of substrate is uniform throughout and equal to the concentration in the effluent.
2. Steady state conditions prevail so that organisms are sheared from the surface of the biofilm at a rate equal to their growth, thereby maintaining a biofilm of constant thickness.
3. The turbulence level in the reactor fluid is sufficient to keep detached organisms in suspension so that they can be carried out in the effluent.
4. Both fixed and detached organisms contribute to substrate removal.
5. Oxygen and other nutrients are present in excess so that the organic substrate is the growth-limiting material.
6. The thickness of the liquid film is uniform over the aerated sector of the discs. Bintanja et al. (7) have shown this to be true.
7. The substrate concentration in the liquid film on the aerated sector depends only on the circumferential angle θ and not upon radial position.
8. When a point on the discs moves from the aerated sector back into the liquid in the tank, the liquid film adhering to the sides is stripped off and completely mixed with the bulk liquid. Zeevalkink et al. (8) have presented evidence that this is not valid. However, due to the complexities which would be introduced if this assumption were not invoked and the lack of information on just how the liquid film actually behaves, it will be used herein.
9. The diameter of the undulations on the surface of a disc is large in comparison to the thickness of the active biofilm so that the overall effectiveness factor developed in the previous section for a flat biofilm is applicable.

Even though the effectiveness factor relationships shown in Figure 8 may be used in modeling the RBC, the mass transfer coefficients employed must reflect the special characteristics of RBC's. As shown in Figure 9, each disc may be divided into two sectors: submerged and aerated. Because the biomass is attached to the disc, it moves through the bulk fluid in the submerged sector, thereby making the external mass transfer coefficient in that sector, k_{Ls}, dependent on the rotational speed, ω. Furthermore, as a point on the surface of a disc leaves the submerged sector and enters the aerated sector a thin film of liquid adheres to it and is carried along with it. Although this film may be assumed to have no motion relative to the biofilm on the disc, its thickness is a function of the rotational speed, ω, thereby making the external mass transfer coefficient in that sector, k_{La}, a function of the speed as well.

Mass transfer from a fluid in laminar flow to the surface of a rotating disc was analyzed by von Karman and given by Levich (9):

$$k_{Ls} = e\ (\omega/r_o)^{0.5} \tag{28}$$

where k_{Ls} is the external mass transfer coefficient in cm/sec, e is a proportionality constant which incorporates the fluid properties, ω is the angular velocity of the rotating disc in revolutions/sec, and r_o is the radius of the disc in cm. Although Eq. 28 indicates that the mass transfer coefficient will increase with the square root of the angular velocity, in practice both the proportionality constant and the power on the angular velocity may be different due to deviations from the assumptions made in deriving the equation. For example, Paolini et al. (10), estimated the effect of rotational speed on a mass transfer coefficient which measured predominantly external transfer resistance and found it to vary with ω to the first power. Thus at this time it would be best to express Eq. 28 in general terms.

$$k_{Ls} = e\ (\omega/r_o)^{f} \tag{29}$$

with the exponent f an experimentally determined value.

In the aerated sector, the liquid forms a stagnant layer on top of the biofilm. Thus it may be assumed that the mass transfer coefficient is equal to the diffusivity divided by the thickness of the stagnant liquid layer. The thickness of liquid film entrained by a flat plate withdrawn verticaly from a quiescent liquid has been analyzed by Landau and Levich (11) and found to be

$$\delta_L = a\ v^{2/3} \tag{30}$$

where δ_L is the film thickness in cm, V is the withdrawal velocity in cm/sec, and a is a parameter which is dependent upon the fluid properties. Since the withdrawal velocity of a point on a rotating disc depends upon its radial position some average value should be used, suggesting that the average film thickness is given by

$$\delta_L = a_1\ \omega^{2/3} \tag{31}$$

where a_1 is a proportionality constant which depends upon the fluid properties and the size of the disc. Bintanja et al. (7) and Zeevalkink et al.(12) have experimentally measured the effect of rotational speed on the thickness of liquid film entrained upon a clean disc and have found it to vary with the square root of the angular velocity. This suggests that the exponent in Eq. 31 is not yet well established and thus should be considered to be an experimentally determined value, a_2. Furthermore, because of surface irregularities in the biofilm which were not present in the studies cited above, it is necessary to add an arbitrary amount to the predicted thickness to account for retention of fluid by the biofilm, as suggested by Famularo, et al. (13). Hence, a more appropriate form might be

$$\delta_L = a_1 \omega^{a_2} + a_3 \qquad (32)$$

Famularo et al. (13) chose a value of 25 μm for a_3. The mass transfer coefficient for the aerated sector may then be estimated by dividing the diffusivity of the solute in water, D, by the film thickness:

$$k_{La} = D/(a_1 \omega^{a_2} + a_3) \qquad (33)$$

Once the external mass transfer coefficients have been estimated for each of the sectors the overall effectiveness factors may be determined for each with Figure 8. They will not be the same. Consequently, we will use η_{oa} and η_{os} to denote the overall effectiveness factors for the aerated and submerged sectors, respectively. The substrate consumption rate per unit area of biofilm in each sector is given by Eq. 24 with the appropriate value for the overall effectiveness factor.

Returning to Figure 9 we may now develop the mass balance equations which constitute the model. Taking the liquid volume in the trough as the control volume, V, we see that substrate is brought in by two streams; the influent flow, F, and the liquid film on the aerated sectors of the discs, F_L. Two streams also comprise the output; the effluent flow, F, and the liquid film entrained by the rotating discs, F_L. Both the organisms in suspension and the organisms on the submerged sector of the discs consume substrate. Thus, the steady state mass balance equation for substrate in the trough is

$$FS_o + F_L S_{LR} - FS_b - F_L S_b - A_s\,[q_m S_b/(K_s + S_b)]\,\eta_{os} - [\mu_m\, S_b/(K_s + S_b)] X_b V/Y = 0 \qquad (34)$$

where S_o, S_b, and S_{LR} are the substrate concentrations in the influent, trough (and effluent), and entrained fluid returning to the trough, respectively, A_s is the total wetted surface area of biofilm in the submerged sector; X_b is the concentration of suspended microorganisms in the submerged sector; μ_m is their maximum specific growth rate; and Y is their observed yield. Since little is known about death, decay, and true growth yield in fixed-film reactors, the concentration of cells in suspension is best estimated from the observed yield and assumption 2, which stated that cells are sheared from the biofilm at a rate equal to their growth. Thus, if X_o, the concentration of cells in the influent, is zero, X_b is given by

$$X_b = Y\,(S_o - S_b) \qquad (35)$$

Substitution of Eq. 35 for X_b and rearrangement yields

$$\frac{1}{\tau} = \left(\frac{A_s}{V}\right) \frac{[q_m\ S_b/(K_s + S_b)]}{(S_o - S_b)}\ \eta_{os} + [\mu_m\ S_b/(K_s + S_b)] + \left(\frac{F_L}{V}\right) \frac{(S_b - S_{LR})}{(S_o - S_b)} \tag{36}$$

The volumetric flow rate of the entrained liquid is given by

$$F_L = 2N\pi(r_o^2 - r_i^2)\ \delta_L\omega \tag{37}$$

where N is the number of discs in the reactor. If we ignore the edges of the discs the total wetted surface area in the submerged sector, A_s, is

$$A_s = N(2\pi r_o^2)\ \{\cos^{-1}(r_i/r_o)/\pi - (r_i/r_o)\ [1 - (r_i/r_o)^2]^{0.5}/\pi\} \tag{38}$$

where $2\pi r_o^2 N$ is the total surface area of the discs, A_s, and the expression in braces, { }, represents the fractional submergence of the discs, f_s. Equation 37 and 38 may be substituted into Eq. 36 making the performance of the RBC a function of the number, size, degree of submergence and speed of rotation of the discs.

Equation 36 also contains a term for the concentration of substrate in the liquid film being returned to the trough by the rotation of the discs. Thus before it can be solved, information must be available about that concentration. As the disc rotates through the aerated sector the microorganisms in the biofilm remove substrate from the entrained liquid thereby making S_{LR} less than S_b. Furthermore, since the entrained film acts like a plug flow reactor, S_{LR} will depend upon S_b. This dependency can be shown by writing a steady state mass balance equation for substrate in a differential element of the entrained film as shown in Figure 9. When performing this balance assumption 7 is invoked, i.e. that the substrate concentration at any point in the aerated sector of the disc, S_L, is a function of only θ and is independent of the radial position of the point. Furthermore, because the volume of liquid film is small with respect to the biofilm area the contribution of the suspended organisms to substrate removal will be small in comparison to the contribution by the biofilm. Thus it can be ignored, leading to

$$(F_L/2N)S_L\Big|_{\theta} - (F_L/2N)S_L\Big|_{\theta+\Delta\theta} - [q_m\ S_L/(K_s + S_L)]\eta_{oa}\ \pi(r_o^2 - r_i^2)\ \Delta\theta/2\pi = 0 \tag{39}$$

Substituting Eq. 33 for F_L and taking the limit leads to

$$\frac{dS_L}{d\theta} = \frac{-[q_m\ S_L/(K_s + S_L)]\ \eta_{oa}}{2\pi\delta_L\omega} \tag{40}$$

Because of assumption 7 the boundary conditions may be approximated by

$$S_L = S_b \quad \text{at} \quad \theta = 0 \tag{41}$$

$$S_L = S_{LR} \quad \text{at} \quad \theta = \theta_A \tag{42}$$

Equations 36 and 40 are the mass balance equations which describe the removal of substrate by an RBC and they must be solved simultaneously. For a given set of reactor conditions the unknowns are S_{LR}, S_b, and η_{os} in Eq. 36 and S_b and η_{oa} in Eq. 40. As we saw earlier, however, η_{oa} and η_{os} are functions of S_b as well as of the physical and operational characteristics of the reactor as given by Figure 8. This prevents Eq. 40 from being integrated analytically and requires that a numerical solution be employed. Furthermore, because Eq. 36 is a nonlinear algebraic equation a numerical technique is applicable to it as well. It must be recognized, however, that these two equations are coupled by the boundary conditions, Eqs. 41 and 42. The easiest way to solve the equations for S_b is to use an iterative approach on S_b. A value of S_b is assumed and used to estimate η_{oa}. Equation 40 is then solved for S_{LR}. The assumed value for S_b is also used to estimate η_{os}, which is then used along with S_{LR} to calculate a new value of S_b using Eq. 36. The iteration ends when the calculated value of S_b is the same as the assumed one. Under some circumstances it may be desirable to refine the iteration technique to account for changes in η_{oa}. The overall effectiveness factor for the aerated sector, η_{oa}, is actually a function of S_L, and thus could change along the aerated sector as S_L changes. If the change in S_L is large, it may be desirable to subdivide the aerated sector into several segments and use a different value of η_{oa} in each. As long as the change in S_b is relatively small, however, such a refinement will not be necessary.

When the substrate concentration in the liquid trough, S_b, is less than the saturation constant, K_s, the model may be simplified somewhat. This follows from the fact that S_L is smaller than S_b, so that S_L will be much smaller than K_s, allowing the first order approximation of the Monod equation to be used to characterize the rate of substrate removal in the aerated sector. Consequently Eq. 40 reduces to

$$\frac{dS_L}{d\theta} = \frac{-[q_m\, S_L/K_s]\,\eta_{oa}}{2\pi\delta_L\omega} \tag{43}$$

The boundary conditions are the same. Furthermore, for a first order reaction the overall effectiveness factor is independent of the substrate concentration and is given by

$$\eta_{oa} = \frac{1}{q_m/K_s k_{La} + \phi/\tanh\phi} \tag{44}$$

Consequently Eq. 43 may be integrated directly to give

$$S_{LR} = S_b \exp\{-q_m\,\eta_{oa}[\cos^{-1}(-r_i/r_o)]/\pi\delta_L\omega K_s\} \tag{45}$$

This in turn may be substituted directly into Eq. 36 giving a single equation to be solved to determine the effect of operational parameters upon RBC performance:

$$\frac{1}{\tau} = \left(\frac{A_s}{V}\right)\frac{[q_m\,S_b/(K_s + S_b)]}{S_o - S_b}\,\eta_{os} + \mu_m\,S_b/(K_s + S_b) + \left(\frac{F_L}{V}\right)\frac{S_b\,\{1 - \exp\,[-q_m\,\eta_{oa}\,[\cos^{-1}(r_i/r_o)]/\pi\delta_L\omega K_s]\}}{S_o - S_b}. \tag{46}$$

For a given set of reactor conditions the only two unknowns in Eq. 46 are S_b and η_{os} (which is a function of S_b). Thus it still must be solved by an iterative procedure, although a simpler one.

Performance of a Single RBC

The utility of a conceptual model is that it allows one to see how changes in various individual parameters are likely to influence the performance of a reactor. Often the objective is to gain a qualitative understanding of reactor performance which can serve as a guide in research or to application of the reactor in practice. Consequently, predictions of the performance of a single RBC were obtained by numerically solving Eq. 46 using the parameter values given in Table 1. In order to allow the computations to be performed by computer, the relationships in Figure 8 were approximated by the equations given for η_{os} in the Table. The overall effectiveness factor in the aerated sector, η_{oa}, was calculated with Eq. 44 using the equation for k_{La} given in the Table. The values of F_L and A_s were obtained with Eqs. 37 and 38, respectively. All of the values in Table 1 were chosen to represent reasonable estimates available in the literature. No attempt was made to "curve-fit" the equation to the data of a particular application. Nevertheless, the qualitative similarity of the performance curves to those in the literature attests to the general validity of the model.

The effect of flow rate upon the removal of substrate in an RBC containing a fixed number of discs rotating at a fixed speed is shown in Figure 10. A faster flow rate brings a greater mass of substrate into the reactor per unit time, necessitating an increase in the removal rate per unit area of biofilm. The only way that can be accomplished, however, is by increasing the substrate concentration surrounding the biofilm. Consequently, an increase in flow rate requires the effluent substrate concentration to increase. Because the hydraulic loading, i.e. the volume of flow per unit area of biofilm, is commonly used to characterize RBC performance, it is shown across the top of the Figure.

The effects of influent substrate concentration are also shown in Figure 10. Examination of the curves shows that the effluent substrate concentration will increase as the influent concentration is increased. Again, an increase in the influent substrate concentration imposes an increased mass flow rate of substrate upon the reactor. Because the surface area of biomass is fixed, the only way the rate of substrate removal can be increased is by increasing the concentration around the biofilm, thereby increasing both the rate of mass transport and the reaction rate. It is interesting to note that the percentage substrate removal is relatively independent of the influent substrate concentration at low flow rates but is dependent upon it at higher ones. This is due to the fact that at low flow rates the effluent substrate concentrations are low relative to K_s so that the Monod kinetics reduce to first order kinetics for which percentage removal is independent of influent substrate concentration. At higher flow rates the effluent substrate concentrations are significant in comparison to the saturation constant so that Monod kinetics prevail and a limiting substrate removal rate is reached. Consequently, the percentage substrate removal decreases as the influent substrate concentration is increased. This effect would have been greater had the first order approximation not been employed for substrate removal in the aerated sector.

There has been some interest in the field in the use of organic loading (mass per time per unit area of biofilm) as a design parameter. Comparisons of combinations of influent flow and concentration which have the same product, however, shows that a constant organic loading does not result in either a constant effluent substrate concentration or a constant percent removal. This follows from the fact that flow and concentration influence the system in fundamentally different ways. Friedman et al. (14) have arrived at this conclusion experimentally.

Figure 11 shows that as the rotational speed is increased the percentage substrate removal increases up to an upper limit characterized by the other system parameters. Several factors interact to cause this response. First, mass transfer in the bulk liquid increases as ω is increased. Consequently the overall effectiveness factor in the submerged sector is increased, which indicates that the substrate is consumed more rapidly by the submerged biofilm as the speed is increased. The events occurring in the aerated sector are more complicated, however. First, an increase in ω will cause the thickness of the liquid film on the aerated sector to increase, which will make the mass transfer coefficient in the liquid film decrease, thereby decreasing the overall effectiveness factor in the aerated sector. This coupled with the increase in the rotational speed and the thickness of the liquid film, will cause an increase in the concentration of substrate returning to the bulk fluid from the face of the discs. The increase in the thickness of the liquid film, on the other hand, causes the volume of fluid carried with the discs to increase, so that even though the reduction in substrate concentration across the aerated region decreases as ω is increased, the mass of substrate removed per unit time increases slightly. Hence the rate of substrate removal by both sectors increases as the rotational speed is increased, with the result that the effluent substrate concentration will decrease as ω is increased. Figure 11 shows that the effect is more significant when the rotational speed is small. Similar results have been observed by Paolini et al. (10). It should be noted that very high rotational speed could cause excessive shear of the biomass from the discs, thereby disrupting performance. This possible effect is not reflected in the model.

The effect of the number of discs is shown in Figure 12. An increase in the number of discs causes a corresponding increase in both the submerged biofilm area and the volume of fluid carried with the biofilm into the aerated sector. Consequently, the substrate consumption rates in both sectors will increase with an increase in N. Thus, for a given influent substrate concentration an increase in N will result in an increase in the substrate removal rate and a decrease in the effluent substrate concentration.

Figure 13 shows the effect of the fractional submergence of the discs upon the performance of an RBC with a fixed number of discs of fixed size rotating at a constant speed. By reducing the inner radius, r_I, the discs may be submerged to any fraction up to 0.5. An increase in the fractional submergence allows more organisms to grow on a disc, thereby causing more substrate removal in the submerged sector. It also increases the volume of fluid carried with the rotating discs into the aerated sector, thereby causing the substrate removal rate in that sector to increase as well. Consequently the percent substrate removal will increase as the degree of submergence of the discs is increased.

Model for Substrate Removal in a Chain of RBC's

In most situations a chain of RBC's will be used to economically achieve a desired effluent substrate concentration. Such a system may be modeled simply by applying the model for a single RBC to each reactor in turn, starting with the first and working downstream. The output from the first reactor becomes the input to the second, etc. Unless sedimentation is performed between reactors, a significant number of microorganisms will enter each reactor after the first. Their effect may be incorporated by adding another term to the right side of Eq. 46 making it read

$$\frac{1}{\tau} = \left(\frac{A_s}{V}\right) \frac{[q_m S_b/(K_s + S_b)]\, \eta_{os}}{S_o - S_b} + \left(\frac{\mu_m S_b}{K_s + S_b}\right)\left[1 + \frac{X_o}{Y(S_o - S_b)}\right] + \left(\frac{F_L}{V}\right) \frac{S_b \{1 - \exp [-q_m\, \eta_{oa} [\cos^{-1}(-r_i/r_o)]/\pi\delta_L\, \omega K_s]\}}{S_o - S_b} \tag{47}$$

where X_o is the concentration of cells entering the stage being modeled. It is just equal to the observed yield times the substrate removed in the preceding stage plus the cells entering the preceding stage. Note that the effectiveness factors must be determined for each stage to account for the effects of the substrate concentration and the physical characteristics of each stage.

Performance of a Chain of RBC's

Predictions of the performance of a chain of RBC's were obtained using the approach outlined in the preceding section with the parameter values in Table 1. Since the effect of such factors as rotational speed, degree of submergence, etc. are the same for the chain that they were for a single reactor this discussion will be limited to the effects of the influent characteristics on the substrate removal profile through a six reactor chain. For computational convenience the simplified model was employed but it should be recognized that the assumption of first order kinetics in the aerated sector will be weakest in the first one or two reactors in the chain.

Figure 14 shows the substrate concentration profile along the chain plotted as a semi-log function of stage number. Even though the concentration changes in discrete jumps, a continuous curve has been drawn for illustrative purposes. Examination of that figure shows that the plots are not linear, but are concave downward. This is a direct result of the fact that substrate removal rate obeys a saturation function, i.e. an upper limit is reached as the substrate concentration increases. Such behavior has been observed in practice, particularly at higher substrate concentrations (15). Furthermore, in the latter stages, where the substrate concentration is lowest, the plots become almost linear, suggesting that substrate removal is behaving in a first-order manner. This too has been observed in practice (15). Although not demonstrated in the Figure, the linearity of a plot will depend upon the values of the parameters describing both microbial kinetics and mass transport. Thus linearity of semi-log plots should not be considered to be a fundamental characteristic of RBC's.

The effect of hydraulic loading on the substrate concentration profile is shown in Figure 15 where the hydraulic loading has been changed by changing the number of discs in each reactor. Thus, the greater the number of discs, the lower the hydraulic loading. As would be expected, an increase in hydraulic loading requires more substrate to be removed by the later stages in the chain. More significant, however, is the fact that the linearity of the plot is best at lower hydraulic loading where the substrate concentrations are low.

EXTENSION OF THE MODEL TO TWO LIMITING SUBSTRATES

One assumption made in the development of the conceptual model was that all nutrients except the carbon and energy source were present in excess. This is sometimes a poor assumption, however, particularly when the influent substrate concentration is high because under those conditions oxygen can become rate-limiting as well (14). Consequently, in order for the model to accurately reflect the performance of the process under all conditions, the situation of two limiting nutrients should be considered. Although we have not yet carried out simulations of this situation, we felt that it was important to discuss how it could be incorporated into the conceptual model because by so doing several needs for further research will become apparent.

Carbon - Oxygen Limitation

In order to incorporate the effects of oxygen concentration several things are needed: rate equations which express the removal rate of the organic substrate in terms of the concentrations of it (S) and of the dissolved oxygen (O) in the liquid; rate equations which express the removal rate of dissolved oxygen in terms of the same concentrations; and rate equations for the transfer of oxygen from the atmosphere to the liquid. The latter two sets of rate equations will be used to write additional mass balance equations in each sector (for oxygen) which must be solved along with the ones already incorporated into the model.

Consider first the rate equation for the removal of organic substrate by the suspended biomass. This may be written as a double Monod equation (16):

$$-r_s = \frac{\mu_m X_b}{Y}\left(\frac{S_b}{K_s + S_b}\right)\left(\frac{O_b}{K_o + O_b}\right) \qquad (48)$$

where O_b is the dissolved oxygen concentration in the bulk fluid and K_o is a saturation constant for dissolved oxygen.

The rate equation for the removal of organic matter by the biofilm can be approached in a similar manner but the effects of mass transport must be considered since it determines the concentrations of substrate and dissolved oxygen surrounding the organisms. By analogy with the single substrate model, this can be done by applying an overall effectiveness factor for each material to the reaction rate expression based upon the concentration of each material in the bulk fluid:

$$-r_s = q_m\left(\frac{S_b \ \eta_{oS}}{K_s + S_b}\right)\left(\frac{O_b \ \eta_{oO}}{K_o + O_b}\right) \qquad (49)$$

where η_{oS} is the overall effectiveness factor for the organic substrate and η_{oO} is the overall effectiveness factor for dissolved oxygen. Furthermore, we saw previously that it was necessary to evaluate the effectiveness factor for the aerated and submerged sectors separately. Consequently, when two rate-limiting materials are considered, four effectiveness factors must be used; η_{oSa}, for the carbon substrate in the aerated sector; η_{oSs}, for the carbon substrate in the submerged sector; η_{oOa}, for dissolved oxygen in the liquid film of the aerated sector; and η_{oOs}, for dissolved oxygen in the submerged sector. Although the same equations may be used to evaluate the mass transfer coefficients for oxygen and substrate in a given sector, it should be recognized that the numerical values will be different because of differences in their diffusivities for the two materials. Thus the four effectiveness factors obtained from Figure 8 will all be different.

The rate equation for the utilization of dissolved oxygen may be approached in a similar manner. Thus the rate equation for the utilization of dissolved oxygen by the suspended microorganisms would be:

$$-r_o = \frac{a'\mu_m X_b}{Y}\left(\frac{S_b}{K_s + S_b}\right)\left(\frac{O_b}{K_o + O_b}\right) \tag{50}$$

where a' is a stoichiometric conversion factor relating substrate removal to oxygen uptake. Because Y is an observed yield, and not a true growth yield, this expression takes into account oxygen uptake associated with cell decay as well as that coupled to substrate oxidation. The equation for oxygen utilization by the biofilm is similar to Eq. 49:

$$-r_o = a'q_m\left(\frac{S_b\ \eta_{oS}}{K_s + S_b}\right)\left(\frac{O_b\ \eta_{oO}}{K_o + O_b}\right) \tag{51}$$

This expression is adequate for modeling steady state performance because the conversion factor a' can be chosen to account for oxygen uptake associated with cell decay as well as with substrate oxidation. If one wanted to model a transient situation, on the other hand, in which the mass of organisms in the biofilm was not in equilibrium with the amount of substrate being removed, then it would be necessary to explicitly account for the oxygen utilization associated with cell decay as done by Famularo, et al. (13). This requires that the mass of microorganisms per unit area of biofilm be known. Unfortunately, little is known about how various physical factors such as rotational speed and substrate loading affect that value. Consequently, this is an area that needs more work before an accurate transient state model can be written. In the mean time, however, the approach outlined above should give reasonable results for the steady state situation.

The final thing that is needed to complete the oxygen-carbon limitation model is a set of rate equations for oxygen transfer to the liquid in each sector. Consider first the aerated sector, where oxygen is transferred directly from air to the liquid film. The usual approach to modeling oxygen tranfer is to assume that the flux of oxygen across the air-liquid interface is proportional to the driving gradient, $O_e - O_b$, where O_e is the concentration of oxygen that would result in the liquid if the liquid were in equilibrium with the gas and O_b is the concentration actually in solution. The proportionality constant is k_{La}', an overall volumetric mass transfer coefficient that takes into account the transport of oxygen across the stagnant layers on either side of the gas-liquid interface. Thus the flux of oxygen into the aerated sector is

$$N_o = k_{La}' \, (O_e - O_b) \tag{52}$$

Famularo, et al. (13) have assumed that the mass transfer coefficient for oxygen into the liquid film in the aerated sector is the same as the mass transfer coefficient for oxygen within that liquid film. This may or may not be true and more research is needed to determine if it is.

Oxygen enters the bulk liquid in the trough from two sources: the return of entrained liquid from the aerated sector and from direct transfer caused by turbulence generated by the rotation of the discs. The rate equation for direct transfer would be similar to Eq. 52 except that the mass transfer coefficient would be for the submerged sector, k_{Ls}', and it must include an interfacial area for transfer. They would be measured together as one parameter analagous to the mass transfer coefficient commonly used for aeration. A number of studies have been performed on oxygen transfer in RBC's (7, 12, 17) but they all combine oxygen transfer to the aerated sector and to the bulk fluid into one term, i.e. they were all done by using the clean water, sodium sulfite, transient state test on RBC's without biofilm. Consequently, the change in dissolved oxygen concentration in the bulk liquid was due to oxygen in the entrained liquid as well as to direct transfer by turbulence. During operation of an RBC, the amount of oxygen transferred to the bulk liquid by the entrained liquid will depend upon the extent of substrate utilization in the aerated sector, and thus the modeling of oxygen transfer to the bulk liquid requires the use of separate terms for direct transfer and transfer from the entrained liquid. This has been recognized by those studying oxygen transfer (8), but to date (Jan. 1980) no papers have appeared which characterize the effects of disc rotational speed, diameter, etc. on k_{Ls}', the oxygen transfer coefficient by turbulence alone. Until such studies have been done it will be impossible to write a fundamental conceptual model incorporating the effects of dissolved oxygen. Hopefully this situation will be remedied in the near future so that the approach outlined herein can be utilized.

Once the appropriate rate equations have been written, extension of the model to the carbon-oxygen limited situation becomes simply a matter of adding two additional mass balance equations to the ones listed previously. Since the substrate and oxygen mass balances are coupled through stoichiometry (the parameter a') and kinetics (the double Monod equation) they must be solved simultaneously. Because the oxygen concentration in the aerated sector is likely to vary over a broad range, it may be necessary to divide that sector into several segments, as done by Famularo, et al. (13) and to solve them sequentially in an iterative fashion. While the solution technique may be involved, it is relatively straight-forward and should present no difficulties.

Carbon-Nitrate Limitation

There has recently been considerable interest in the use of RBC's for denitrification. In this situation, nitrate-nitrogen serves as the terminal electron acceptor for oxidation of an externally supplied carbon source and electron donor. The modeling of this situation can also be handled by a dual substrate limitation model analagous to that for oxygen (1). Thus, denitrification can be handled in the same way as that described in the previous section. The only difference will be that the nitrate will be supplied in the influent so that it will not be necessary to have expressions for nitrate mass transfer to the liquid. In other words, the equations for denitrification represent a simplification of the general case for carbon-oxygen limitation.

SUMMARY

A conceptual model has been presented for an RBC which incorporates the effects of biological kinetics and physical mass transfer. This was done by utilizing the effectiveness factor approach in which the reaction rate expression is written in terms of the bulk substrate concentration and is then corrected for the effects of mass transfer by multiplying by the effectiveness factor. Two effectiveness factors are required: one for the submerged sector of the reactor and one for the aerated sector. Each factor includes terms for mass transport up to the biofilm as well as diffusion within the biofilm. Figure 8 allows determination of the effectiveness factor for any physical situation through the use of dimensionless groups. The model was utilized to show the effects of influent flow rate and concentration, disc rotational speed, and degree of submergence upon the performance of a single RBC as well as RBC's in series. Finally, a technique was described whereby the conceptual model could be extended to include the effects of oxygen transfer and dissolved oxygen concentration.

ACKNOWLEDGEMENT

Portions of this material were taken from reference (1). The figures herein are reproduced with permission of Marcel Dekker, Inc., New York. The authors appreciate the comments of Margaret Kupferle concerning the carbon-oxygen limited model.

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TABLE 1

NOMINAL PARAMETER VALUES USED FOR FIGURES 10-15

Symbol	Value
a_s	$100\ cm^2/cm^3$
D_e	$0.02\ cm^2/hr = 5.556 \times 10^{-6}\ cm^2/sec$
F	$300\ l/hr = 8.333 \times 10^2\ cm^3/sec$
k_{Ls}	$0.1863\ (\omega/r_o)^{1/2}$ cm/sec
k_{La}	D_e/δ_L cm/sec
L_f	0.01 cm
N	44
q_m	$1\ mg/(cm^2)(hr) = 2.778 \times 10^{-7}\ g/(cm^2)(sec)$
r_o	150 cm
r_i	30 cm
S_o	$100\ mg/l = 1 \times 10^{-4}\ g/cm^3$
V	3000 liters $= 3 \times 10^6\ cm^3$
Y	0.5 mg cells/mg substrate
δ_L	$2.144 \times 10^{-3}\ [\omega(r_o + r_i)/2]^{2/3}$ cm
μ_m	$0.2\ hr^{-1} = 5.556 \times 10^{-5}\ sec^{-1}$
K_s	50 mg/l substrate
ω	2 rpm = 2/60 revolutions/sec
η_{os} at $\psi = 5$	$0.1987\phi_f^{-1.869} + 0.832\phi_f^{-1.3681}$; $1.65<\phi_f<10$
	$0.1987\phi_f^{-1.869} + 0.7045\phi_f^{-1.0362}$; $1.25<\phi_f<1.65$
$\psi = 1$	$0.66\phi_f^{-1.864}$
$\psi = 10$	$1.1904\phi_f^{-1.3681}$

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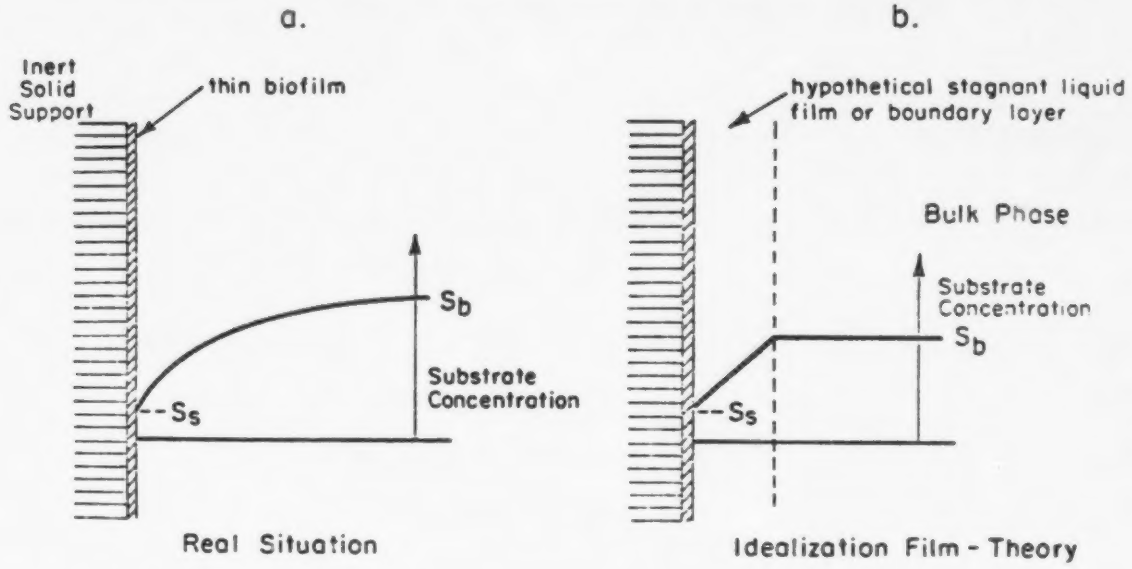
a.
b.
Inert Solid Support
thin biofilm
S_b
Substrate Concentration
S_s
Real Situation
hypothetical stagnant liquid film or boundary layer
Bulk Phase
Substrate Concentration
S_b
S_s
Idealization Film - Theory

Figure 1 - Actual and idealized substrate concentration profiles for reaction at a surface with external mass transfer resistance.

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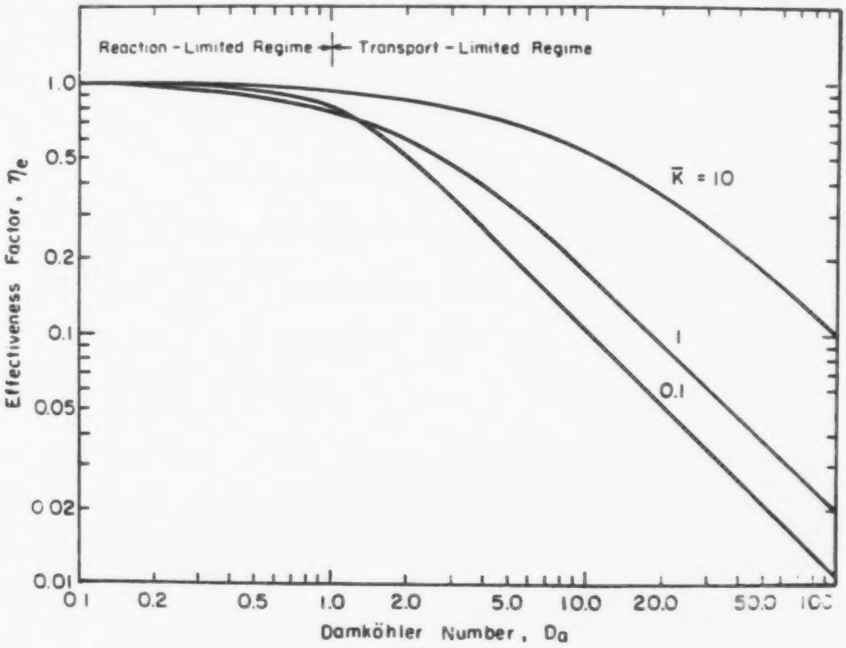

Figure 2 - Effect of Damköhler number on the effectiveness factor for external mass transfer resistance.

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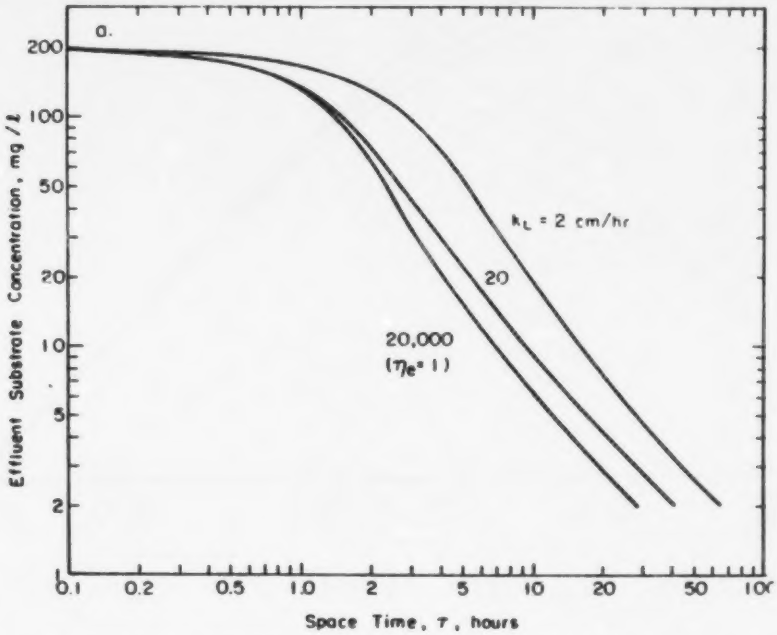

Figure 3 - Effects of external mass transfer resistance on the removal of substrate by a CSTR containing a biofilm

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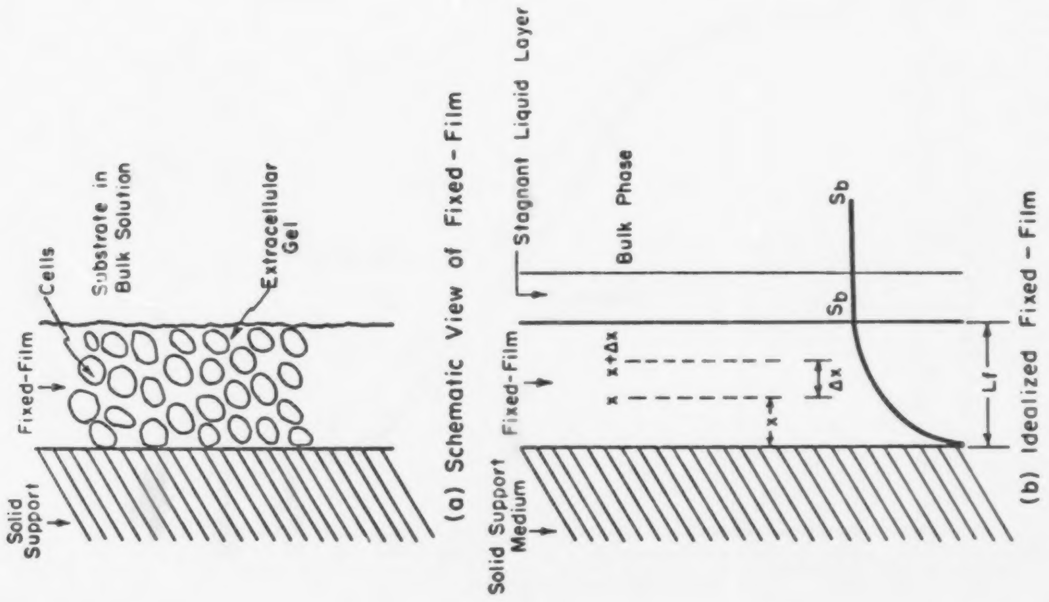
Solid Support
Fixed-Film
Cells
Substrate in Bulk Solution
Extracellular Gel
(a) Schematic View of Fixed-Film
Solid Support Medium
Fixed-Film
Stagnant Liquid Layer
Bulk Phase
S_b
S_b
x
x+Δx
Δx
x
Lf
(b) Idealized Fixed-Film

Figure 4 - Substrate concentration profiles for reaction within a thick biofilm.

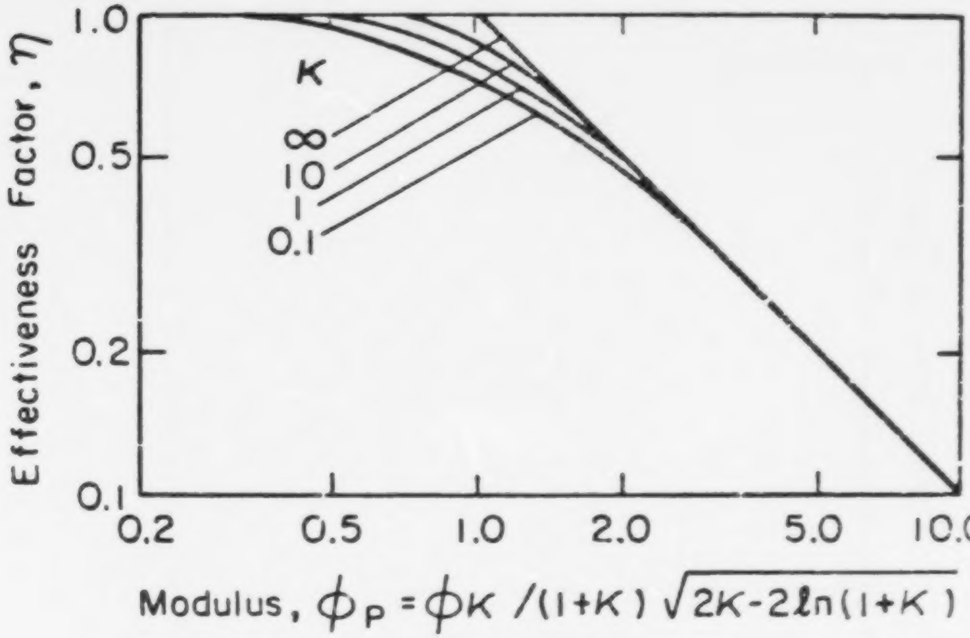

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Figure 5 - Internal effectiveness factor for Monod kinetics within a flat biofilm (After Pitcher (4)).

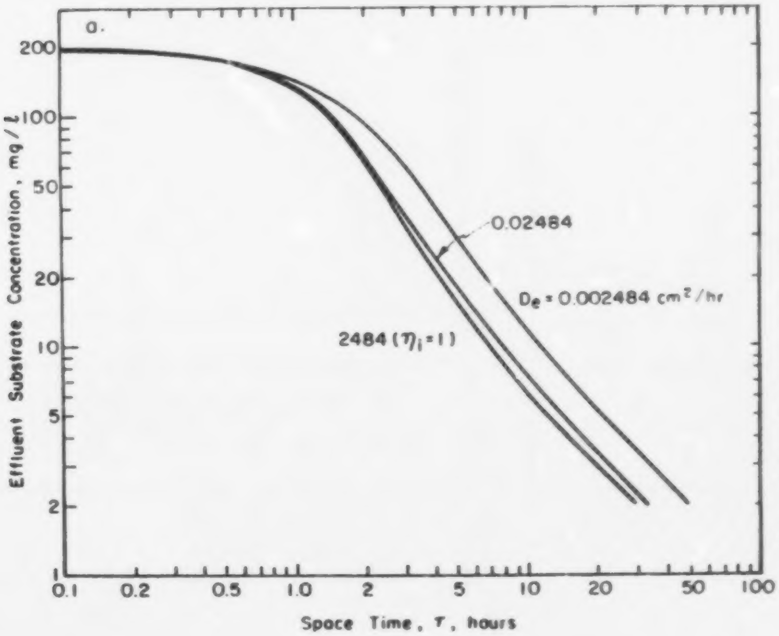

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Figure 6 - Effect of internal mass transfer resistance on the removal of substrate by a CSTR containing a biofilm.

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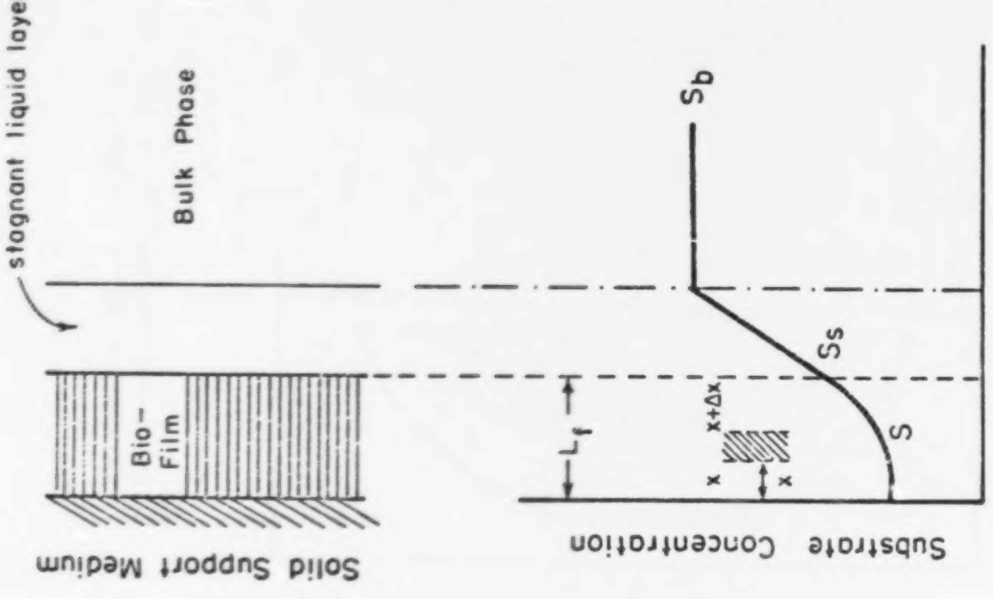
stagnant liquid layer
Bulk Phase
Bio-Film
Solid Support Medium
L_f
x
x
x+Δx
S
Ss
Sb
Substrate Concentration
Distance

Figure 7 - Substrate concentration profiles for reaction within a thick biofilm with external mass transfer resistance.

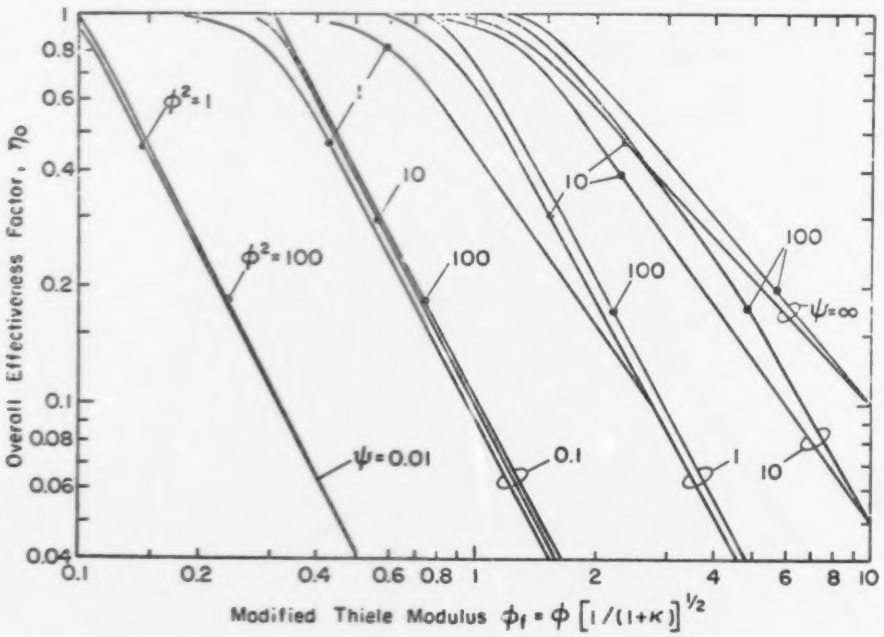

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Figure 8 - Overall effectiveness factor for Monod kinetics with a flat biofilm with external mass transfer resistance (After Fink et al.(6)).

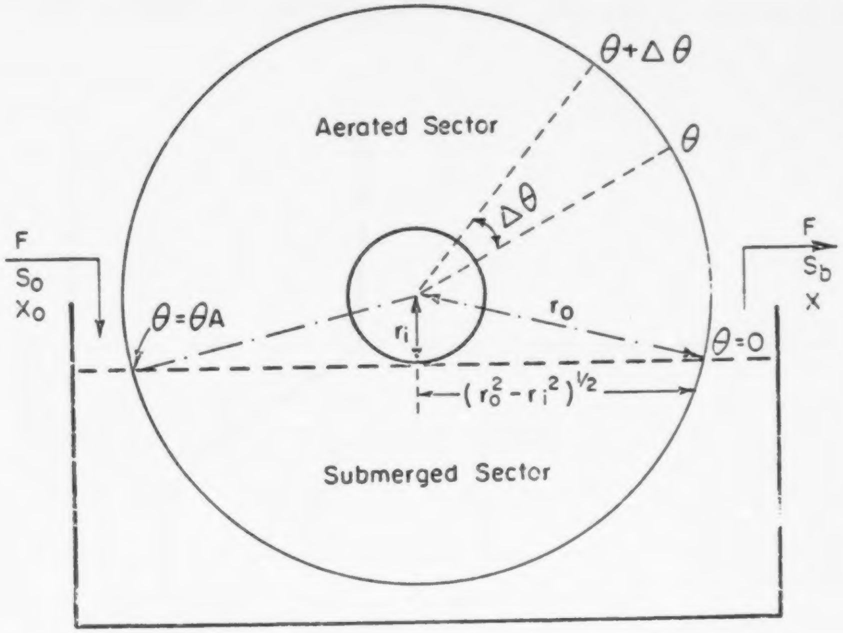
θ+Δθ
Aerated Sector
θ
Δθ
F
S0
X0
θ=θA
ri
r0
θ=0
(r0² - ri²)^1/2
Submerged Sector
F
Sb
X

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Figure 9 - Schematic diagram of a single disc in an RBC.

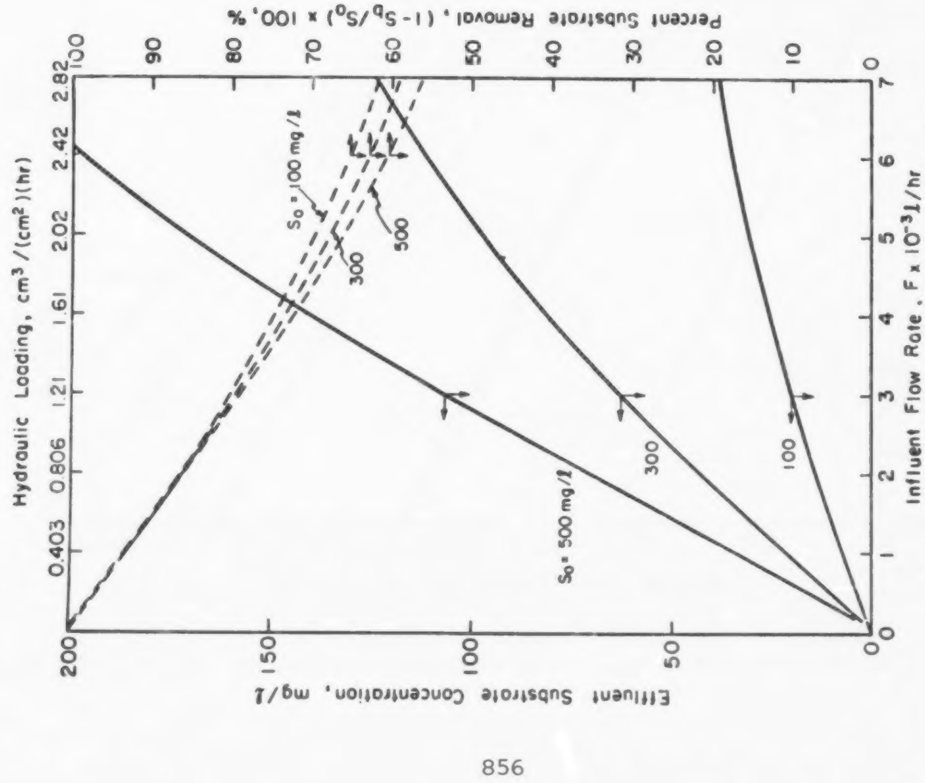

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Figure 10 - Effects of influent flow rate and influent substrate concentration on the performance of a single RBC.

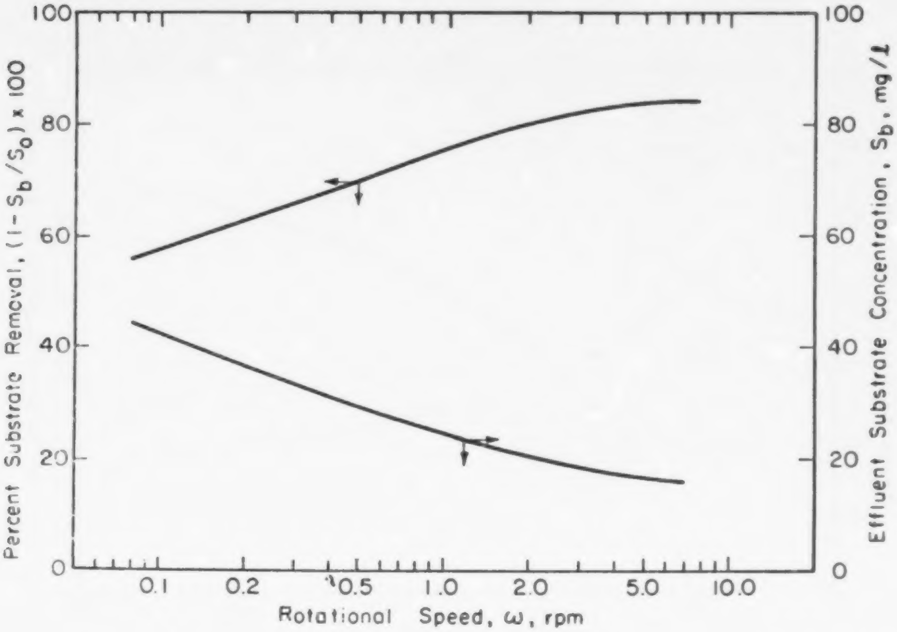

Figure 11 - Effect of rotational speed on the performance of a single RBC.

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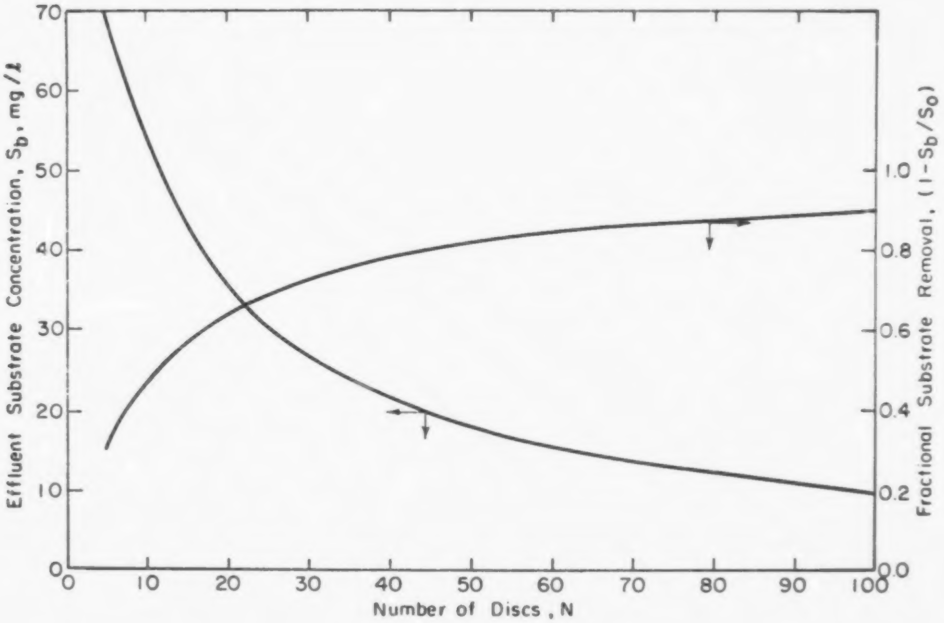

Figure 12 - Effect of the number of discs on the performance of a single RBC.

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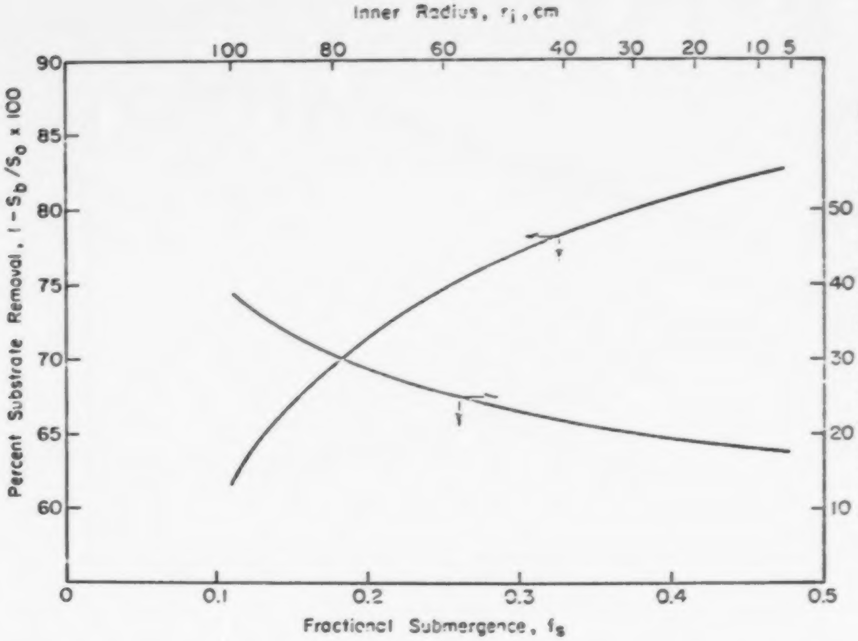

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Figure 13 - Effect of the degree of submergence of the discs on the performance of a single RBC.

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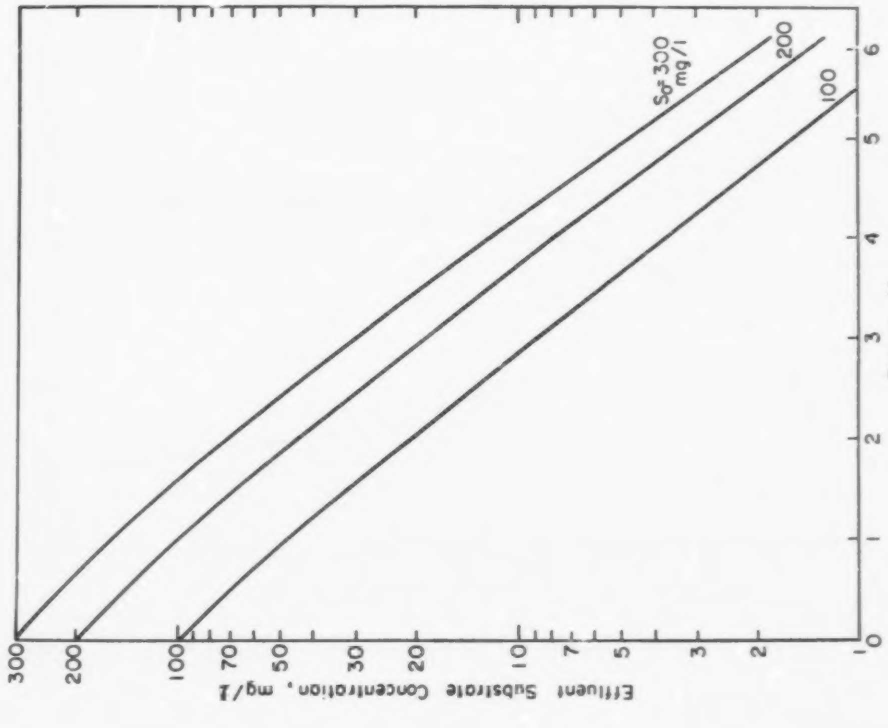

Figure 14 - Effect of influent substrate concentration on the substrate concentration profile through a series of six RBC's.

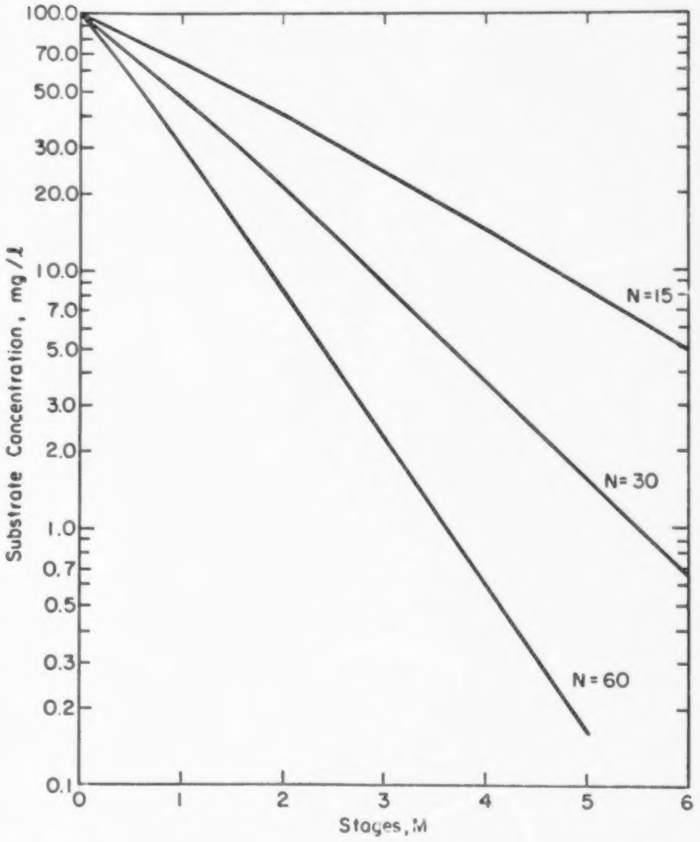

Figure 15 - Effect of hydraulic loading (F/N) on the substrate concentration profile through a series of six RBC's.

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RECARBONATION OF WASTEWATER USING THE ROTATING BIOLOGICAL CONTACTOR

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INTRODUCTION

Carbon, as carbon dioxide, is readily available through atmospheric cycles and from natural carbonate alkalinity for assimilation by organisms present in receiving waters. Nitrogen can be drawn from the atmosphere by nitrogen-fixing bacteria, or is available in various forms from surface runoff. However, phosphorus is not available through atmospheric cycles and it is further limited by its sedimentation cycle. For this reason, phosphorus most frequently limits excessive eutrophication, and is the most controllable nutrient found in man's wastewaters.[1,2]

Chemical and biological wastewater treatment processes can result in significant reductions of phosphorus. Chemical-biological removal is accomplished with the same chemicals used in purely chemical-physical removal.

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The chemicals can be added to raw, degritted sewage; to the primary effluent; to the aeration tank or trickling filter; to the secondary clarifier influent or effluent. The choice of chemical and point of addition depend on effluent standards, alkalinity, pH, sludge dewatering, sludge disposal, chemical costs and chemical availability.

Low-level lime addition has recently been applied prior to activated sludge aeration basins with favorable results. Activated sludge aeration basins contain enough buffering capacity due to microbial activity to reduce the elevated pH to a reasonable level for biological treatment, normally below pH 8.5. The buffering capacity of rotating biological contactors has not been considered for pH adjustment. Consequently, lime addition has not been considered for pH adjustment, nor for use in wastewater treatment systems, except as a tertiary process employing higher pH levels and recarbonation equipment.

The RBC process was selected for evaluation of its ability to provide recarbonation of high pH wastewaters (pH 9.5) and to nitrify these same wastewaters. The choice of this treatment scheme considered the relatively low energy and operating costs of the RBC process, simplicity of operation, and flexibility with respect to the upgrading of existing treatment plants.

This paper describes phosphorus removal by low-level lime (pH 9.5) precipitation in the primary clarifier, followed by biological recarbonation using the RBC. Of equal importance in this research effort was to show the enhancement of nitrification by the chemically induced increase in pH. Goals were set to provide criteria for upgrading existing army wastewater treatment facilities to meet NPDES permit limitations, as opposed to research directed toward completely replacing such facilities.

LITERATURE REVIEW

Phosphorus Removal

Phosphorus removal encompasses several technologies. Biological, chemical-physical, chemical-biological and physical techniques have successfully removed phosphorus from wastewaters. Biological techniques employ microorganisms to remove phosphorus through synthesis, metabolic processes and adsorption incorporating the phosphorus into biological solids. Chemical-physical techniques utilize precipitation, coagulation, flocculation, adsorption, sedimentation and filtration to incorporate the phosphorus in a chemical sludge. Common chemicals used include lime, iron salts, alum, sodium aluminate and polyelectrolytes. Rare earth elements have also successfully precipitated phosphorus. All of these techniques involve the conversion of soluble phosphorus to a solid form amenable to conventional solid-liquid separation technology.

High-Lime Treatment

High pH (lime) treatment of wastewaters is usually associated with tertiary processes for phosphorus or ammonia removal. These processes operate at a pH value of approximately pH 11.0 and therefore result in increased sludge production. At pH 11.0 magnesium precipitation is essentially complete. Formation of hydroxyapatite $Ca_5OH(PO_4)_3$ and tricalcium phosphate $Ca_3PO_4)_2$ begins at pH 9.0 and precipitation proceeds toward completion with approximately 80 percent removal occurring at pH 9.5 to pH 10.0. The increased dosage of lime required to achieve a pH value of 11.0 not only increases the amount of chemical sludge produced, but also creates a requirement for a recarbonation process prior to effluent discharge. This technique implies an obvious increase in construction and operation and maintenance costs.

Low-Level Lime

Low-level lime dosing to achieve a pH 9.5-10.0 can reduce phosphorus to levels to meet the biological system's nutrient demands, without an excessively high pH. Schmid and McKinney[3] used relatively low doses of lime (150 mg/l) in a primary system operating at a pH of 9.5 to achieve 80 percent total phosphorus removals, 60 percent BOD removals and 90 percent reduction of suspended solids. They report that both the orthophosphates and condensed phosphates can be readily precipitated with lime in conjunction with recarbonation in the secondary treatment system by microbial production of CO_2. Their combined system removed 90-95 percent of the total phosphorus in a domestic wastewater containing 40-50 mg/l total phosphorus (as PO_4) with lime doses generally less than 150 mg/l as calcium hydroxide. Miller, et al.[4] and Cavagrano et al[5] have shown the importance of coagulant aids for effective solid-liquid separation of precipitated calcium. Ferric chloride at doses of 5 mg/l (as Fe) appear to be most effective.

Recarbonation

The source of carbon dioxide for lowering effluent pH levels can be liquid CO_2, a by-product of fuel combustion, stack gases from sludge incinerators or recalcining furnaces, or from the biological oxidation of organic matter. All of these recarbonation techniques require the construction, operation and maintenance of additional plant processes, except for biological recarbonation. Carbon dioxide production through biological oxidation of organics is accomplished in the secondary treatment process thereby requiring no new construction.

Low-level lime addition has recently been applied prior to activated sludge aeration basins with favorable results. The point of addition is to the raw, degritted sewage and the pH is elevated to 9.5 to 10.0. Review of the literature yielded no applications or studies of low-level lime addition prior to a RBC. Lime addition has traditionally been associated with higher pH levels and a recarbonation requirement. Activated sludge aeration basins contain enough buffering capabity due to microbial activity to reduce the elevated pH to a reasonable level for biological treatment, normally below pH 8.5. The buffering capacity of rotating biological contactors has not been considered for pH adjustment. Consequently, lime addition has not been previously attempted for use in systems except as a tertiary process employing higher pH levels and a recarbonation requirement.

MATERIALS AND METHODS

Performance of an RBC was evaluated for recarbonation as a result of secondary treatment and nitrogen removal. Pilot studies used domestic wastewater from the Fort Detrick housing area. The wastewater was shredded by a grinder pump and pumped at a rate of 7 gallons per minute into a 250 gallon tank which acted as a grit chamber with a detention time of 25 minutes. The raw, degritted wastewater was pumped to the rapid mix tank where lime was added. The wastewater (with elevated pH) then flowed by gravity to the flocculation basin. The wastewater further flowed by gravity through the primary settling basin, through the rotating biological contactor (RBC), and through a secondary settling basin. A schematic of the wastewater, flow route is shown in Figure 1.

The lime feed system consisted of a lime slurry tank, a rapid mix tank and an automatic pH control system. The lime slurry tank contained a 2 percent slurry of $Ca(OH)_2$ and was continuously mixed. The rapid mix tank had a detention time of about 5 minutes, depending upon flow rate. The pH measurement for automatic control was taken in the flocculation basin. Feedback by the pH controlled the off-on time of the lime slurry feed pump. Ferric chloride feed for effective solid-liquid separation was used as an integral part of the lime feed system. The ferric chloride was fed into the wastewater at the rapid mix tank, prior to flocculation at 5 mg/l (as Fe).

The flocculation basin has a detention time of about 30 minutes. Mixing was provided by a double paddle rotating at 15 rpm. Baffles were used as part of the flocculation basin to obtain effective floc formation. The primary settling basin had a detention time of 2 to 3 hours depending on flow rate.

The Rotating Biological Contactor (RBC) consisted of four compartments in series. The 0.5 meter plastic media disks provided 250 sq ft of surface area for microbial attachment. The disk was rotated through the liquor at 13 rpm with 40 percent of the fixed-film submerged at any point.

Total Organic Carbon measurements were made on a Beckman Model 915 Total Organic Carbon Analyzer. Ammonia-nitrogen concentrations were measured with Orion specific ion electrode. Dissolved Oxygen and BOD determinations were made using a Delta Scientific Model 2110 Dissolved Oxygen Meter and Probe. Chemical Oxygen Demand, Total Kjehdahl Nitrogen, and phosphorus analyses were made using a Technicon AutoAnalyzer II system according to U.S. EPA approved Technicon methods. Filtered samples were filtered through fiberglass filters, except for phosphorus samples which were filtered through 0.45 μm membrane filters. All other analyses were performed according to Standard Methods.[6]

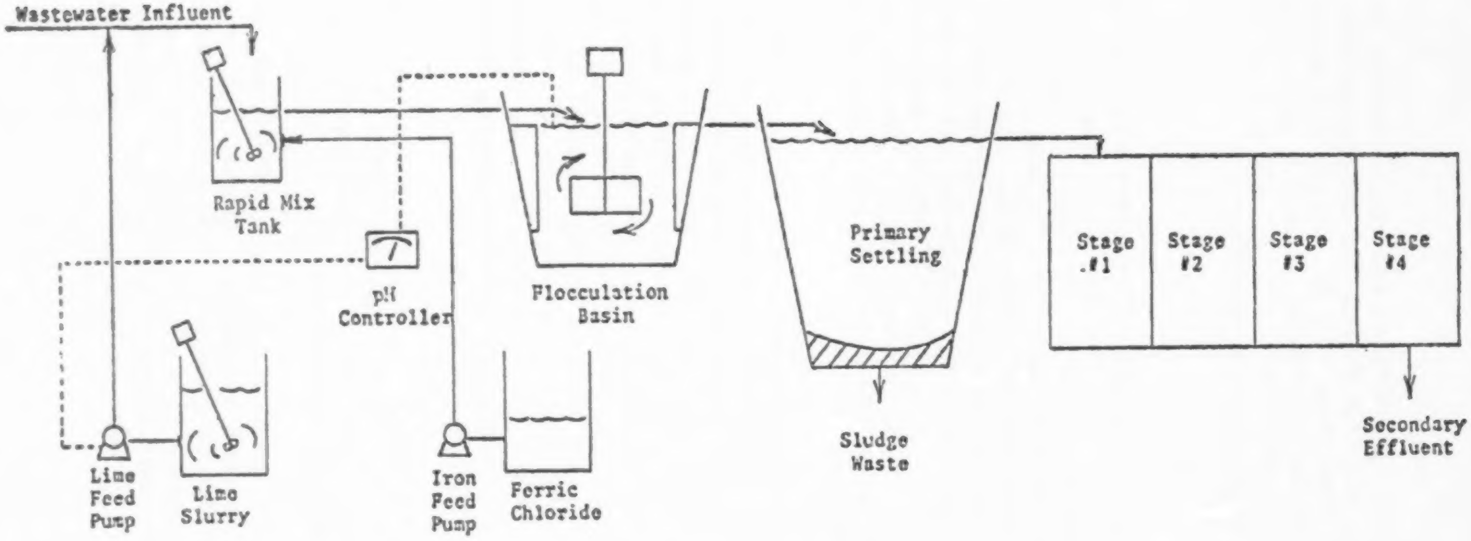
Wastewater Influent
Rapid Mix Tank
pH Controller
Flocculation Basin
Primary Settling
Stage .#1
Stage #2
Stage #3
Stage #4
Lime Feed Pump
Lime Slurry
Iron Feed Pump
Ferric Chloride
Sludge Waste
Secondary Effluent

FIGURE 1. Schematic of the Pilot Scale Lime Addition and Secondary Treatment Processes.

Rotating Biological Contactor

The RBC was evaluated for secondary treatment and recarbonation in conjunction with the low-level lime addition method for phosphorus removal. The RBC was evaluated at pH 9.5 at flow rates of 2.0, 3.0 and 4.0 gpd/sq ft and pH 7.0 at 3.0 gpd/sq ft. The RBC was evaluated at pH 9.5 and 2.5 gpd/sq ft with 30 and 50 percent of the flow feed into Cell #3. All evaluations were accomplished at a wastewater temperature of 15-25°C.

RESULTS AND DISCUSSION

The RBC was initially evaluated for secondary treatment and recarbonation of high pH wastewaters (pH 9.5) derived from low-level lime addition before primary settling for the purpose of phosphorus removal. The hydraulic loading rates tested were 2.0, 3.0, and 4.0 gpd/sq ft. Secondary treatment efficiency was evaluated using filtered and unfiltered BOD_5, and filtered TOC. The degree of treatment received was also determined by monitoring the extent of nitrification at various loading rates. Suspended solids, pH, temperature, and dissolved oxygen levels within the RBC stages were monitored. Sample points included raw wastewater, lime treated primary clarified effluent, and RBC effluent. Limited sampling was conducted within the stages of the RBC to evaluate the progression of treatment.

While low-level lime addition techniques precipitate about 80 percent of the available phosphorus,[3] additional phosphorus is removed through biological assimilation (about 1 percent of the BOD_5 reduction).[7] Figure 2 depects the physical and biological aspects of phosphorus removal. Additional phosphorus removal after secondary treatment is possible by chemical addition for precipitation in the secondary clarifier and filtration. The assimilation of phosphorus was fairly constant irrespective of chemical addition, but the final effluent concentrations were significantly reduced during chemical addition. Phosphorus removal is not the only added benefit of biological recarbonation, but enhancement of nitrification is also achieved (Table 1). The nitrification can be attributed to the greater organics removal before secondary treatment processes and to a more optimal pH for nitrifier activity.

Table 1 shows data supporting the desirability of the biological recarbonation process. While the overall BOD_5 removal was unaffected, the ammonia-nitrogen and phosphorus removals were significantly increased by the low-level lime addition process prior to secondary treatment.

Figure 3 further construes the efficacy of the lime addition process for enhancement of secondary treatment. Figure 3 shows the soluble BOD_5 concentrations with respect to the progression of treatment with and without low-level lime addition at a hydraulic loading rate of 3.0 gpd/sq ft. In each case the primary influent soluble BOD_5 was essentially the same. However, a small portion of the soluble BOD_5 was removed by the low-level lime addition technique for phosphorus removal.

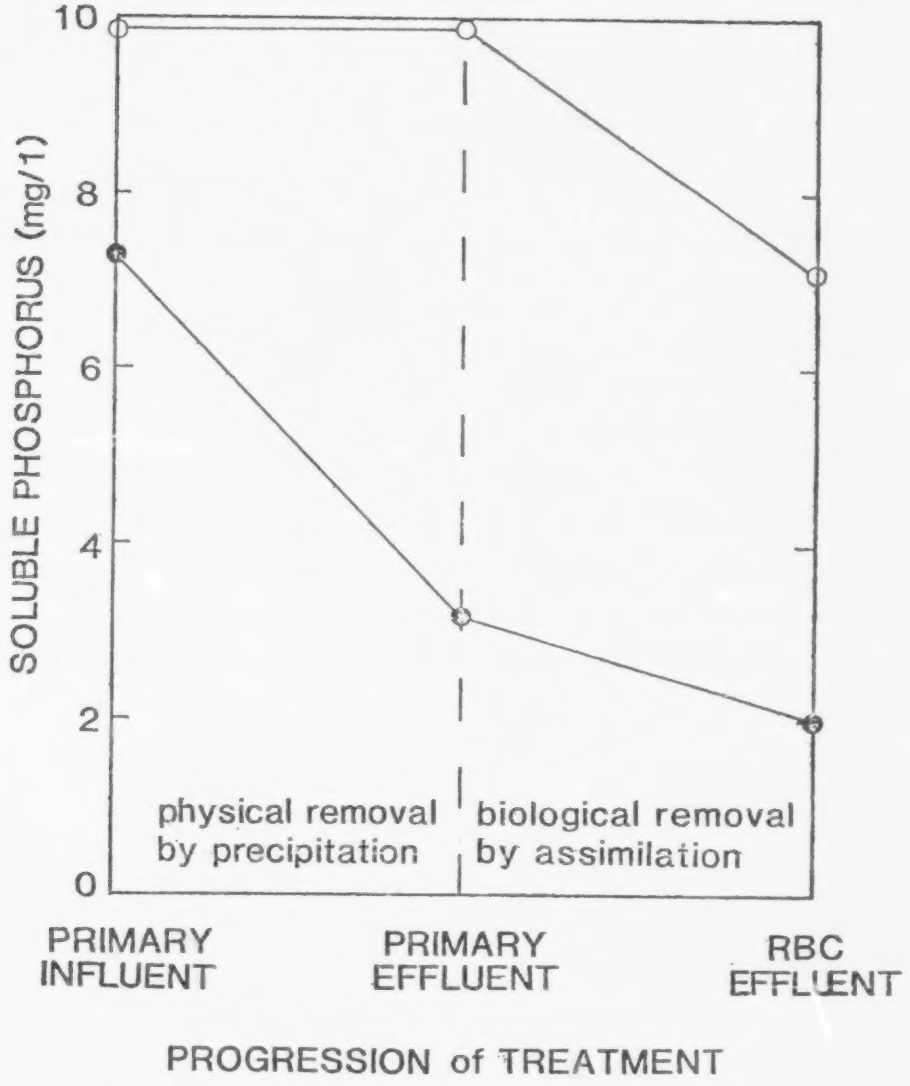

FIGURE 2. Physical and Biological Aspects of Phosphorus Removal are Depicted. Solid circles represent wastewater treated with lime to achieve a pH value of 9.5 followed by primary settling for removal of phosphorus precipitate. Open circles are results from control experiments where no chemical was added to the wastewater. The hydraulic loading rate on the RBC was 3.0 gpd/sq. ft.

TABLE 1. THE EFFECT OF LOW-LEVEL LIME ADDITION ON NUTRIENT REMOVAL. (RBC INFLUENT pH VALUES WERE 9.3 AND 7.1 FOR WASTEWATERS WITH AND WITHOUT LIME ADDITION, RESPECTIVELY. THE HYDRAULIC LOADING RATE WAS 3.0 gpd/sq ft.)

Parameter	Percent Removed	
	With Lime Pretreatment	Without Lime Pretreatment
BOD_5-Unfiltered	88	87
BOD_5-Filtered	92	92
TOC-Filtered	64	45
COD-Filtered	70	71
Phosphorus-Soluble (as P)[a]	37	24
NH_3-N	83	40

[a] Due to assimilation only, does not include chemical precipitation.

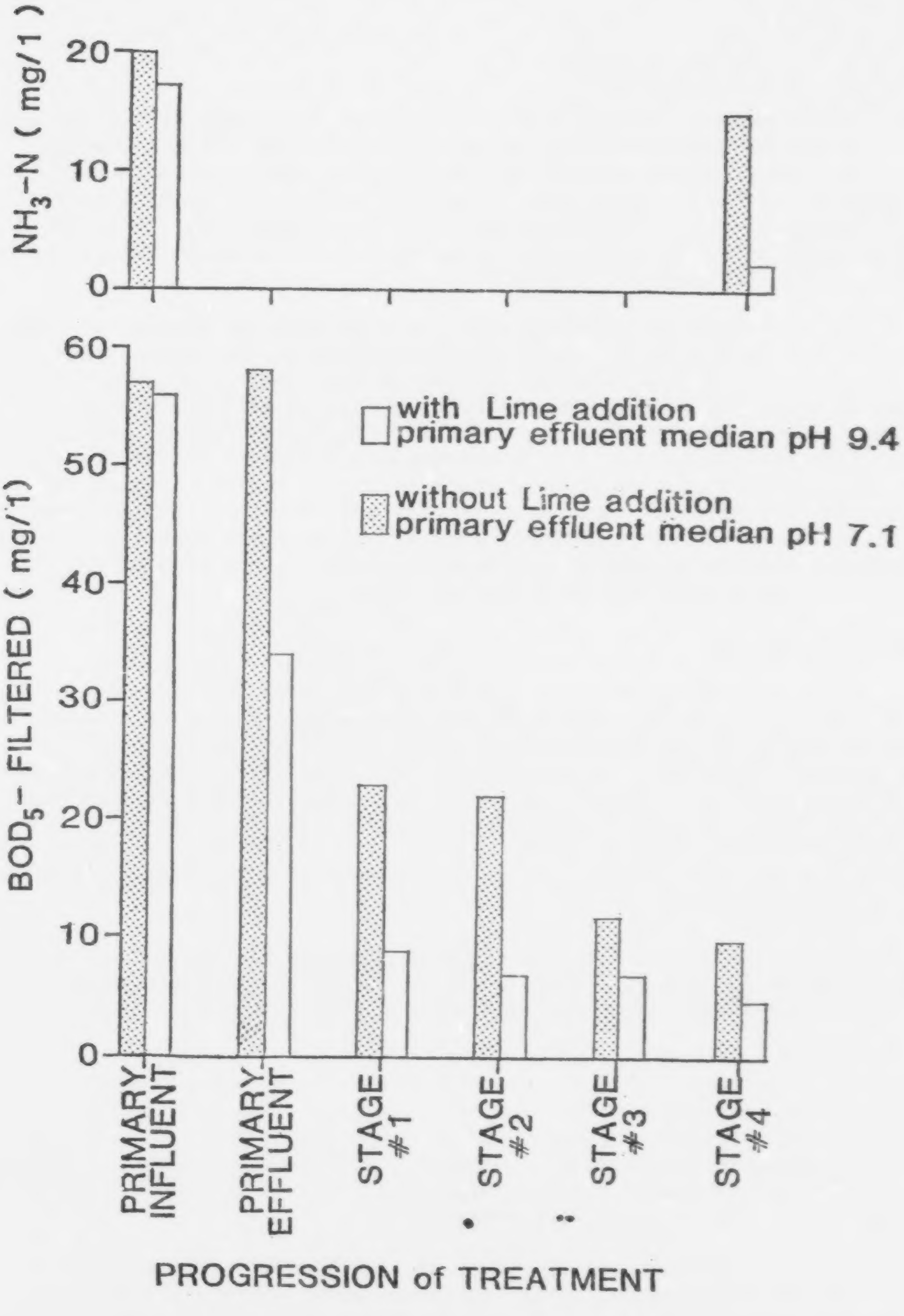

FIGURE 3. Soluble BOD$_5$ Concentrations are Shown With Respect to the Progression of Treatment. Influent and effluent ammonia concentrations are also shown. The hydraulic loading rate was 3.0 gpd/sq. ft. for both conditions.

The difference in BOD_5 reduction by the first stage of the RBC was quite apparent when comparing removal rates with and without low-level lime addition. Eighty percent of the BOD_5 was removed from the high pH wastewater effluent, while only 58 percent was removed without an initial pH adjustment. This indicates that more RBC surface area is available for autotrophic nitrifiers, thereby, allowing for the increased nitrification shown in the top panel of Figure 3.

Nitrification was achieved when the lime addition process for phosphorus removal was incorporated into the wastewater treatment train. Figure 4 illustrates the conversion of ammonia to nitrite plus nitrate. Nitrification is shown relative to the treatment processes with and without the addition of lime. When no lime was added for phosphorus removal and consequent elevation of influent pH, little ammonia was converted to nitrite and nitrate. However, the addition of lime to the influent wastewater allowed for a substantial conversion of ammonia to nitrite and nitrate. The low-level lime addition technique for phosphorus removal not only achieves its primary function, phosphorus removal as shown in Figure 2, but also allows for nitrification without increased surface area.

Hydraulic Loading

Table 2 compares nutrient removal at hydraulic loading rates of 2.0, 3.0, and 4.0 gpd/sq ft. At the higher flow rates (3.0 and 4.0 gpd/sq ft) the wastewater pH was lowered more slowly by the biological degradation of carbonaceous materials. Little variation in phosphorus removal was observed, but ammonia-nitrogen removal decreased dramatically. RBC effluent ammonia-nitrogen levels increased from 2.7 mg/l at 3.0 gpd/sq ft to 18.6 mg/l at a hydraulic loading rate of 4.0 gpd/sq ft. This indicates that nitrification did not occur, most likely as a result of increased RBC surface area requirements for BOD removal.

The percent removal of BOD remained constant as the loading rate was increased to 4.0 gpd/sq ft (Table 3). However, the percent removal of COD, TOC, and NH_3-N declined sharply at the elevated hydraulic loading.

Figure 4 illustrates pH values expressed as a function of the biological treatment received. At 3.0 gpd/sq ft little difference in pH was observed without chemical addition for phosphorus removal. During the period when lime was added to elevate the pH to pH 9.5 for phosphorus removal, the degree of recarbonation decreased with increasing flow rate. Adequate recarbonation and secondary treatment were achieved at all hydraulic loadings tested, however, the degree of nitrification varied.

Figure 5 substantiates the lack of ammonia removal at 4.0 gpd/sq ft as there was no decrease in inorganic carbon, which is indicative of nitrification and concomitant changes in the carbonate buffering system. Also, Figure 5 illustrates the recarbonation of high pH wastewater through increases of inorganic carbon levels. Inorganic carbon levels are expressed as a function of the influent concentration which ranged from 30 to 38 mg/l.

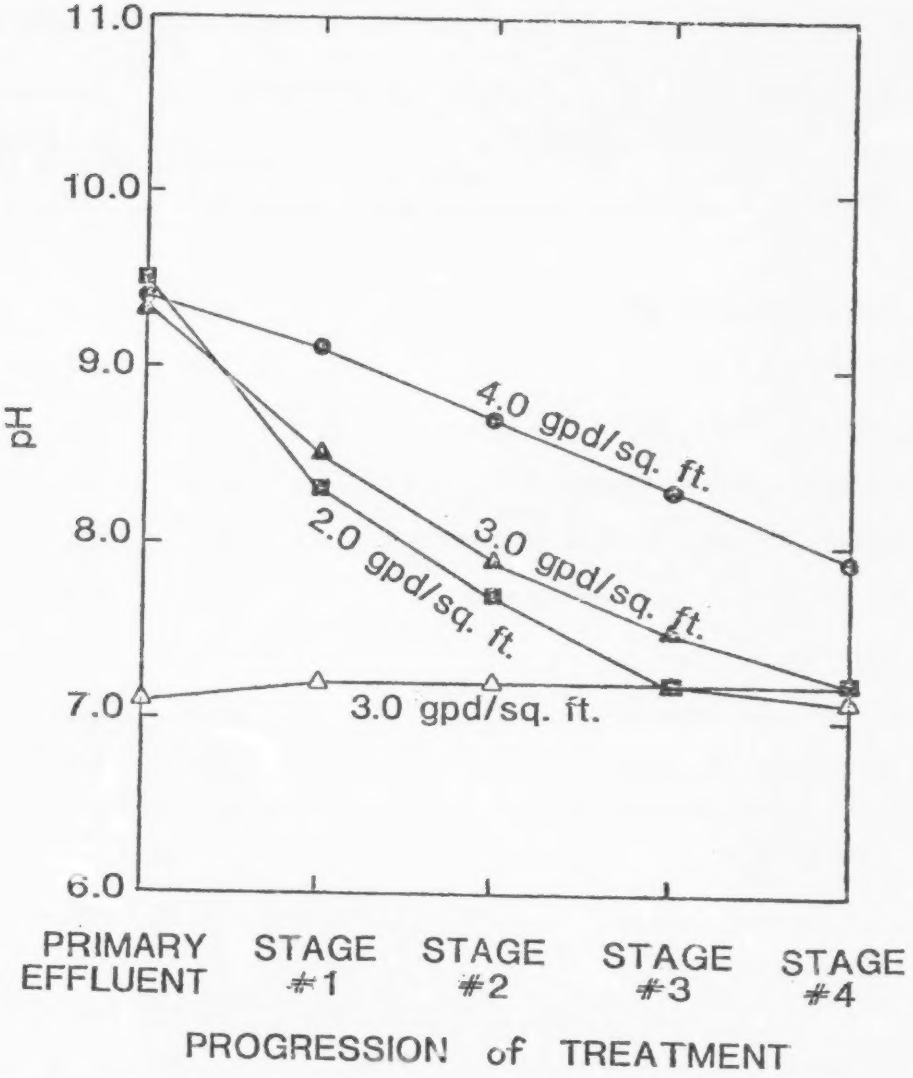

FIGURE 4. The Depicted pH Values are Expressed as a Function of the Biological Treatment Received. Circles, triangles, and squares represent the biological recarbonation effect on wastewater at 2.0, 3.0, and 4.0 gpd/sq. ft., respectively. The solid symbols represent wastewater which was pretreated with lime for phosphorus removal. The open triangles show the constancy of pH values when no lime was added.

TABLE 2. COMPARISON OF NUTRIENT REMOVAL AT THREE HYDRAULIC LOADING RATES

Parameter	Mean Concentrations Hydraulic Loading (gpd/sq ft) 2.0	 3.0	 4.0
pH			
Primary Effluent pH[a]	9.5	9.3	9.4
RBC Effluent pH[a]	7.0	7.2	7.9
Phosphorus-(as P), mg/l			
Influent Soluble Phosphorus	10.5	7.2	9.7
Primary Effluent Soluble Phosphorus	3.8	3.2	3.3
RBC Effluent Soluble Phosphorus	2.3	2.0	2.0
Ammonia-(as N), mg/l			
RBC Influent Ammonia	17.0	15.6	20.9
RBC Effluent Ammonia	2.4	2.7	18.6

a. Median value.

TABLE 3. EFFECT OF HYDRAULIC LOADING RATES ON NUTRIENT REMOVAL. THE RBC INFLUENT pH VALUES WERE pH 9.5, pH 9.3 AND pH 9.4 FOR THE LOADING RATES OF 2.0, 3.0, and 4.0 gpd/sq ft RESPECTIVELY.

Parameter (mg/l)	2.0 gpd/sq ft			3.0 gpd/sq ft			4.0 gpd/sq ft		
	RBC Inf	RBC Eff	% Removed	RBC Inf	RBC Eff	% Removed	RBC Inf	RBC Eff	% Removed
BOD_5(Filtered)	72	2	97	45	4	92	62	6	91
TOC (Filtered)	32	11	66	36	13	64	55	28	49
COD (Filtered)	206	38	81	138	42	70	182	104	43
NH_3-N	17.3	2.4	86	15.6	2.7	83	20.9	18.6	11

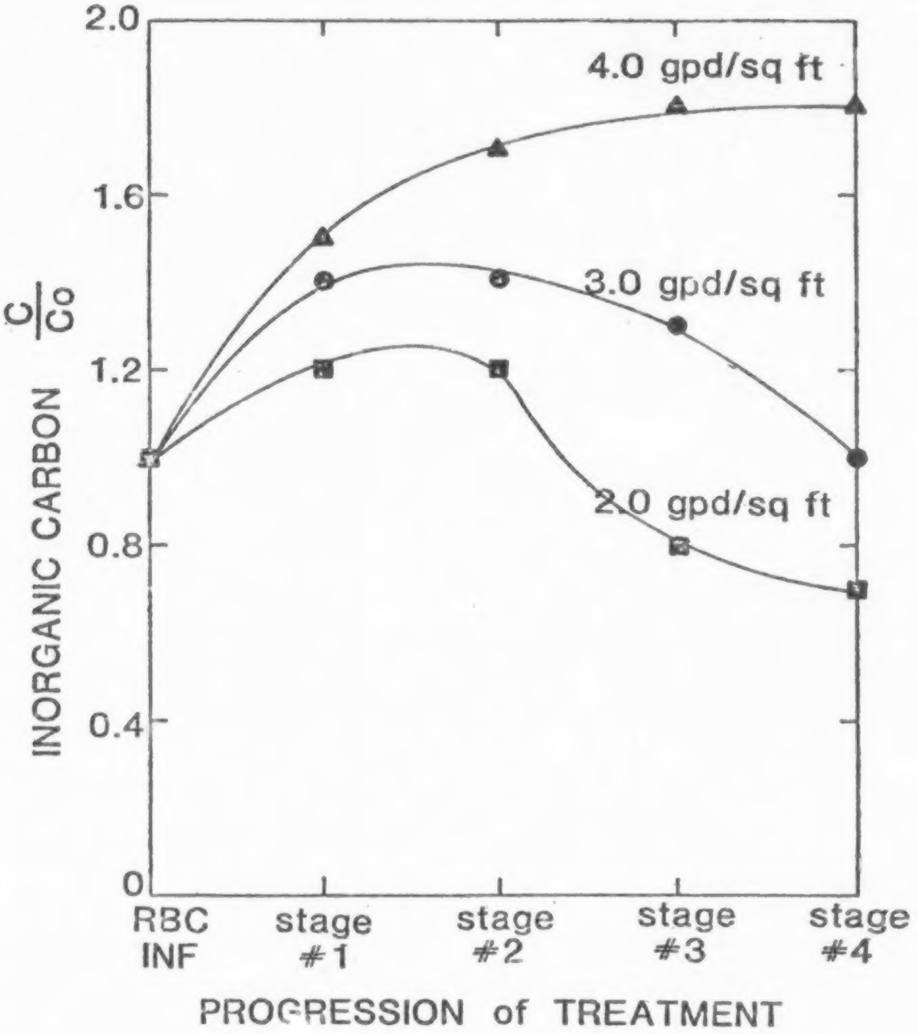

FIGURE 5. Inorganic carbon produced through biological oxidation of organic substrates at various hydraulic loading rates. Inorganic carbon levels are expressed as a function of the influent concentration which ranged from 30 to 38 mg/l.

The production of CO_2 increased with greater hydraulic loading rates and corresponding increases in organic loading. Even though higher levels of inorganic carbon were found at the greater hydraulic loading rates, pH depression proceeded more slowly.

Step-Feed Studies

High pH primary clarified wastewater was applied to the RBC at an overall hydraulic loading rate of 2.5 gpd/sq ft. During control studies, 100 percent of the flow was applied to the first RBC stage and then passed through the successive stages. The two experimental trials diverted 30 percent, and later, 50 percent of the RBC influent to the third stage. Nutrients removed during these tests are shown in Table 4. Large differences in the removal of organic constituents was not observed. The 70/30 split of wastewater influent demonstrated the least degree of treatment, but was attributable to inorganic chemicals appearing in the wastewater as indicated by an increase in COD levels. This elevated COD load did exert an effect on nutrient removal, but did not greatly affect wastewater recarbonation. Table 5 shows the pH depression and TOC removal across the RBC stages.

Figure 6 shows the TOC concentrations in the RBC stages at the various wastewater feed conditions. However, the numbers reported as mg/l are not an accurate measure of removal since more wastewater was applied to the latter RBC stages, but they are representative of the expected effluent concentrations. Figure 7 illustrates TOC removal on a mass per unit area per unit time basis. Even though the 100/0 wastewater application scheme had the lowest initial TOC concentration, it had a relatively high loading rate since none of the flow was diverted to stage number 3 as in the other studies. The addition of primary clarified wastewater to the third RBC stage has resulted in increased TOC removal in the latter stages, but not necessarily an increase in overall removal. As seen in Table 4, the highest percent removal of TOC was observed when the flow was split between stages 1 and 3 yet the lowest effluent concentration was achieved without splitting of the flow. Figure 8 demonstrates the cummulative percent removal of TOC by the three flow application schemes. No large differences were observed in the overall removal for each treatment scheme. However, splitting of the wastewater flow appears to better utilize the available RBC media surface area.

Biologically induced pH depression was achieved with all three loading schemes. The initial reduction of pH in the first two stages was greater with respect to the longer retention times resultant of flow splitting (Figure 9). No adverse effects were observed as a result of introducing high pH wastewater into the third RBC stage. Figures 10 and 11 express recarbonation as a function of the amount of inorganic carbon produced. Figure 10 shows the actual concentration of inorganic carbon present in the RBC stages. Inorganic carbon is produced in stage 1 where much of the BOD and TOC were degraded by microorganisms. As the wastewater pH decreases, aeration continues, and nitrification ensues, the inorganic carbon levels decline across the RBC. When wastewater was applied to the third stage, the production of inorganic carbon again exceeded its removal. The open symbols are expected values if additional wastewater had not been applied to the third RBC stage (Figures 10 and 11).

TABLE 4. NUTRIENT REMOVED BY SECONDARY TREATMENT. THE RBC RECEIVED A HYDRAULIC LOADING OF 2.5 gpd/sq. ft.

	100% to Stage #1			70/30 Split to #1 & #3			50/50 Split to #1 & #3		
	RBC Inf	RBE Eff	% Removed	RBC Inf	RBC Eff	% Removed	RBC Inf	RBC Eff	% Removed
BOD_{Unf}	67	11	84	77	12	84	103	12	88
BOD_{Fil}	45	3	93	53	7	87	64	4	94
TOC	54	26	48	67	38	57	65	29	55
Alkalinity	234	157	-	23	201	-	255	197	-
COD	118	20	83	277	132	52	157	37	83
NH_3-N	14.2	3.5	75	22.2	12.6	43	19.9	7.2	64
NO_2+NO_2	0.10	7.14	-	0.33	6.32	-	0.11	10.19	-
SS	42	11	74	57	37	40	60	13	78
pH[a]	9.2	7.5	-	9.3	7.8	-	9.2	7.6	-

a. Median value.

TABLE 5. TOC and pH VALUES CONCENTRATION WITHIN THE RBC STAGES. THE FLOW WAS SPLIT BETWEEN STAGES #1 and #3 AS INDICATED. THE OVERALL HYDRAULIC LOADING WAS 2.5 gpd/sq. ft.

Parameter	RBC Inf	#1	#2	#3	#4
TOC (mg/l)					
100/0	45	31	24	21	20
70/30	72	42	37	38	35
50/50	68	37	31	36	28
pH[a]					
100/0	9.1	8.9	8.6	8.2	7.9
70/30	9.3	8.8	8.3	8.4	8.0
50/50	9.5	8.7	8.1	8.3	7.7

a. Median value.

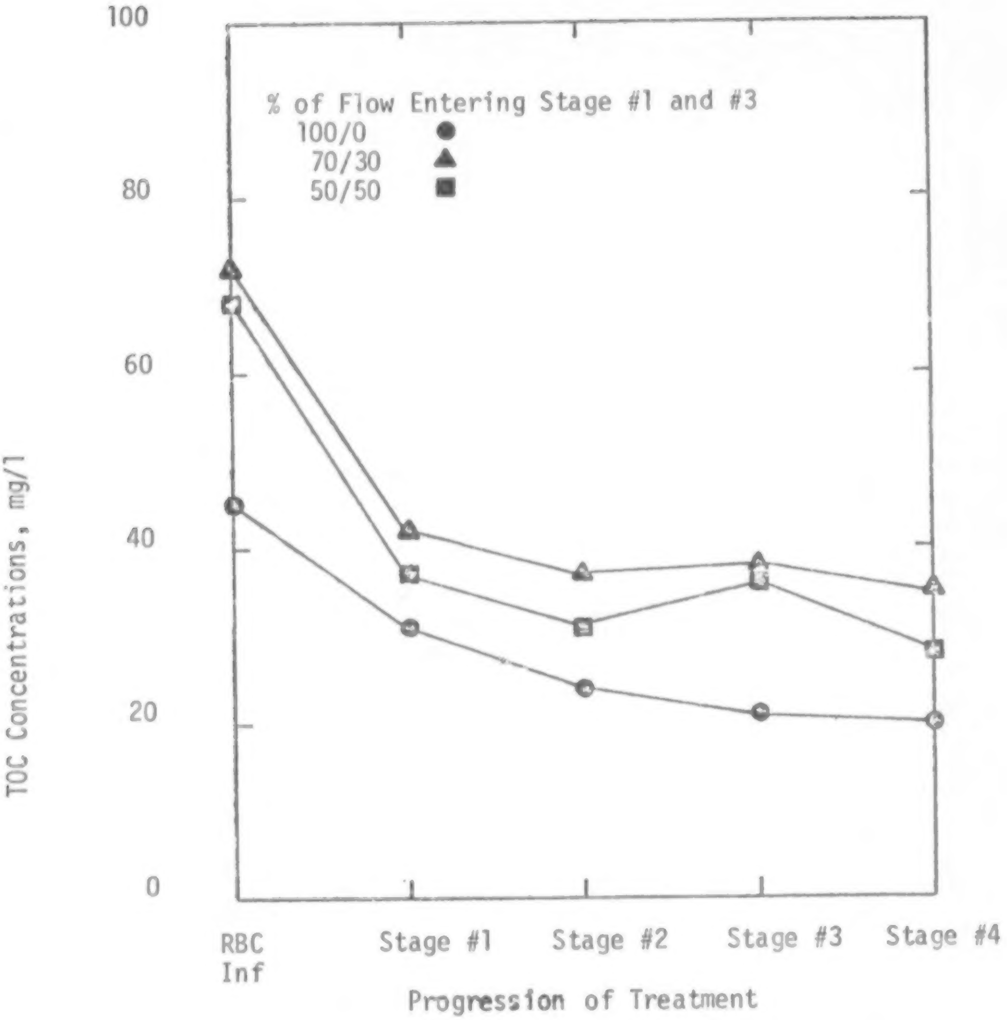

FIGURE 6. TOC Concentrations are Depicted in the RBC Influent and Within the RBC Stages. The Symbols Illustrate the Percent of Flow Entering Stages #1 and #3 as Indicated.

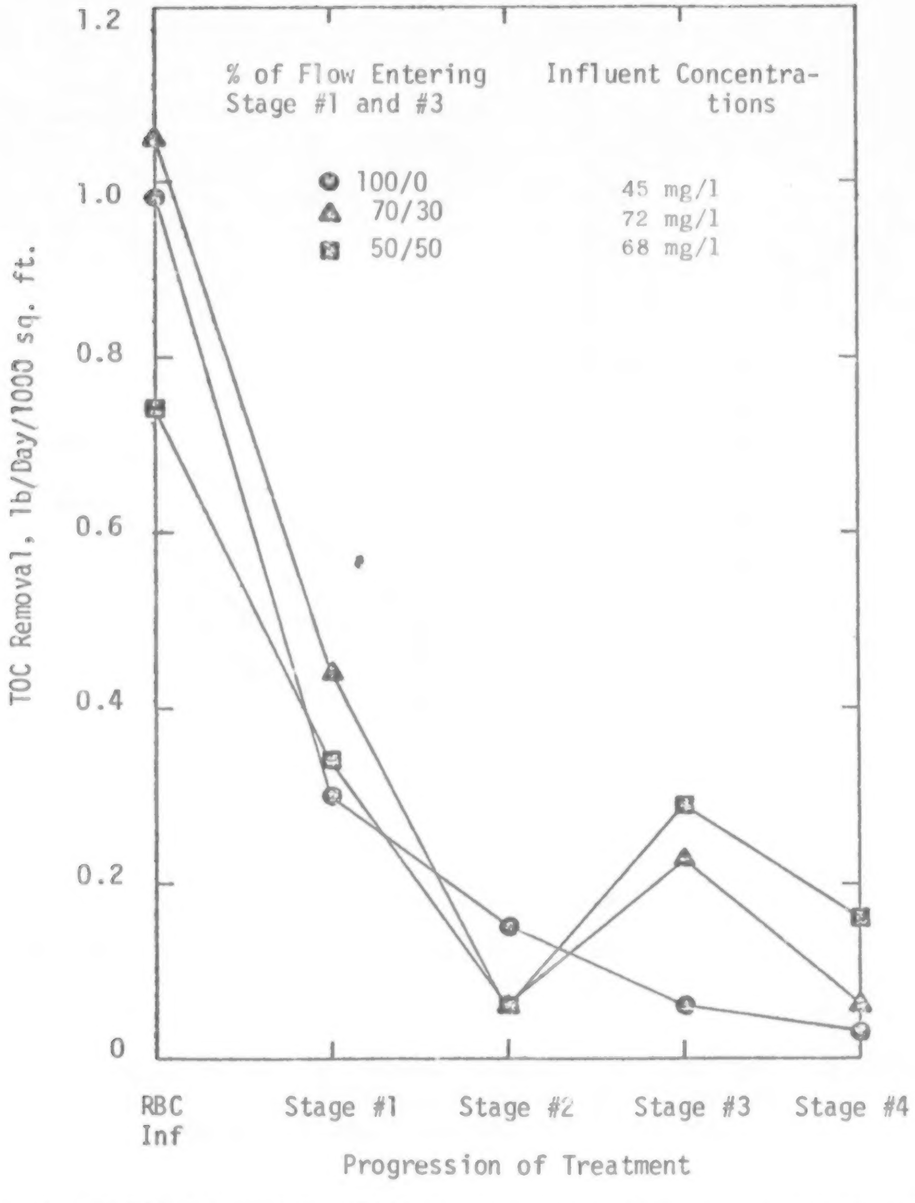

FIGURE 7. Pounds of TOC Removed/Day/1000 sq. ft. are shown for each RBC stage. The symbols illustrate the percent of flow entering Stages #1 and #3.

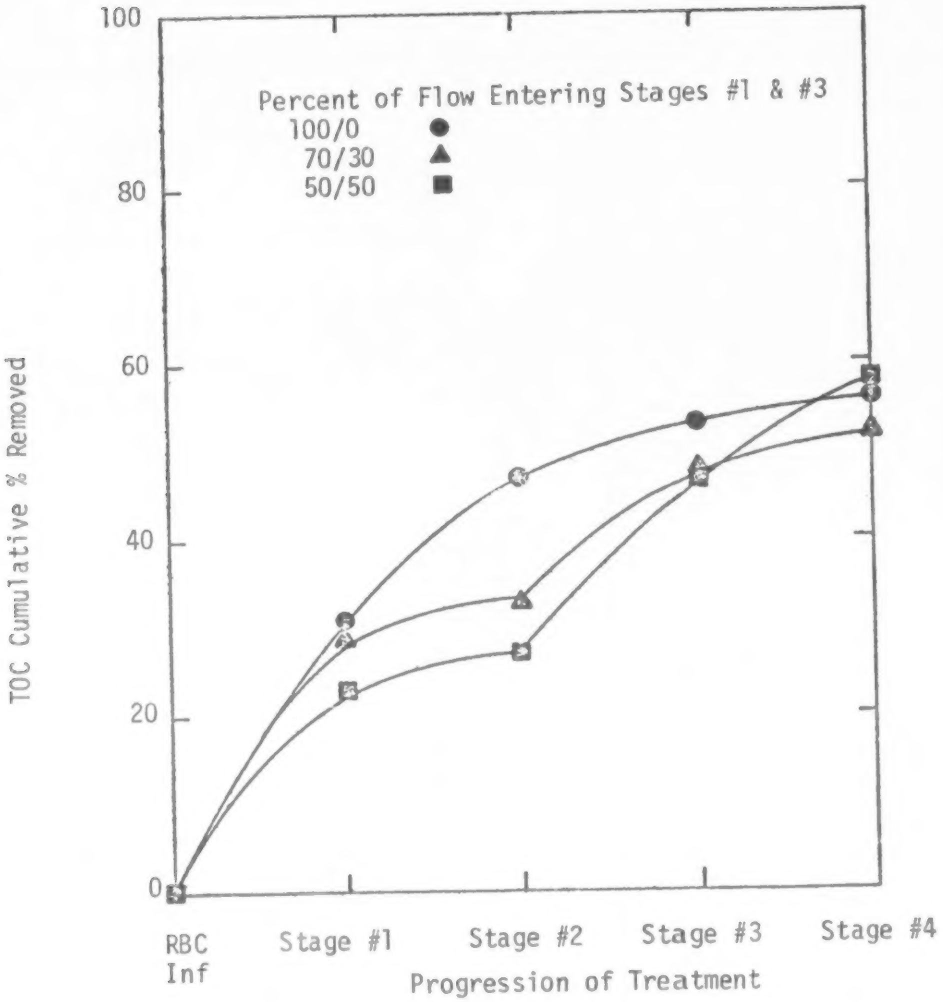

FIGURE 8. The Cumulative Percent Removal of TOC is Shown With Respect to the Degree of Biological Treatment Received. The Symbols Illustrate the Percent of Flow Entering Stages #1 and #3.

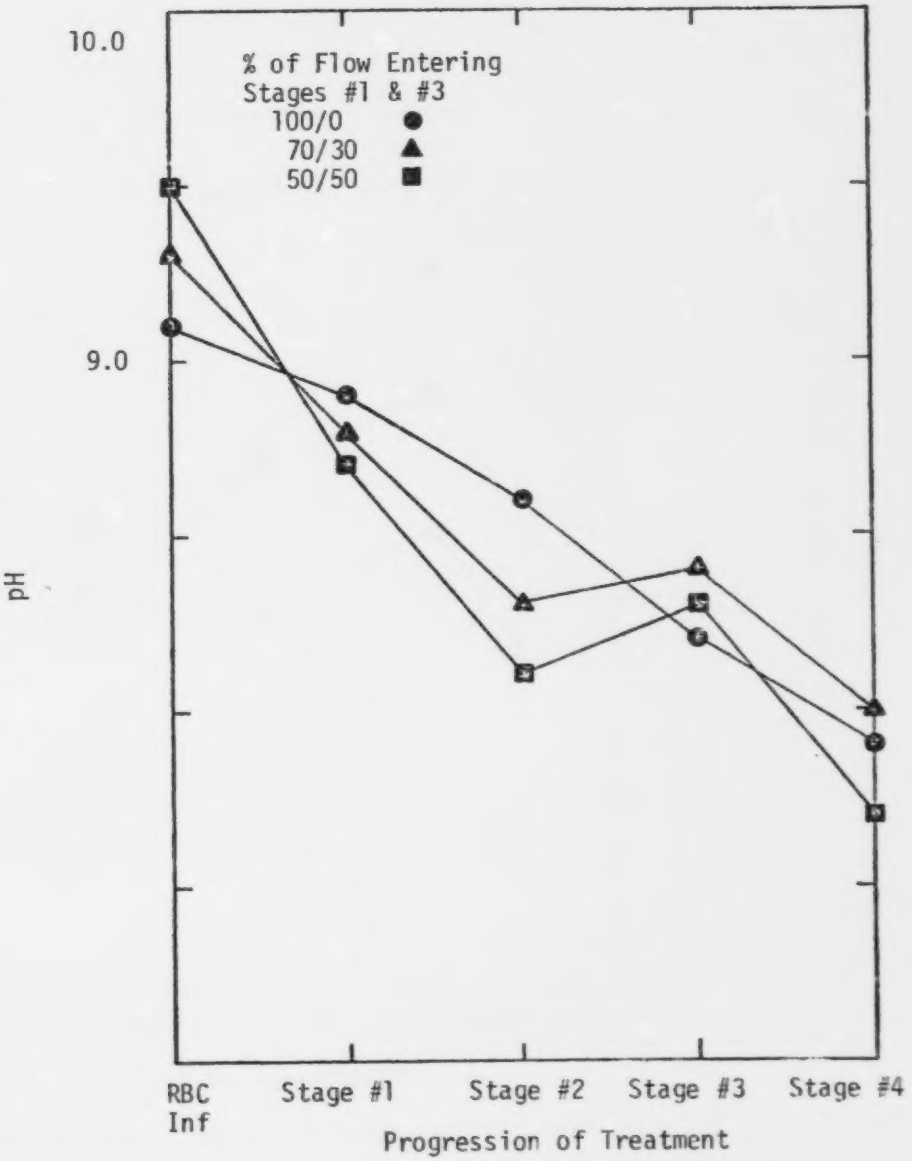

FIGURE 9. Recarbonation is Expressed by the Depression of pH Within Each RBC Stage. The Symbols Illustrate the Percent of Flow Entering Stages #1 and #3.

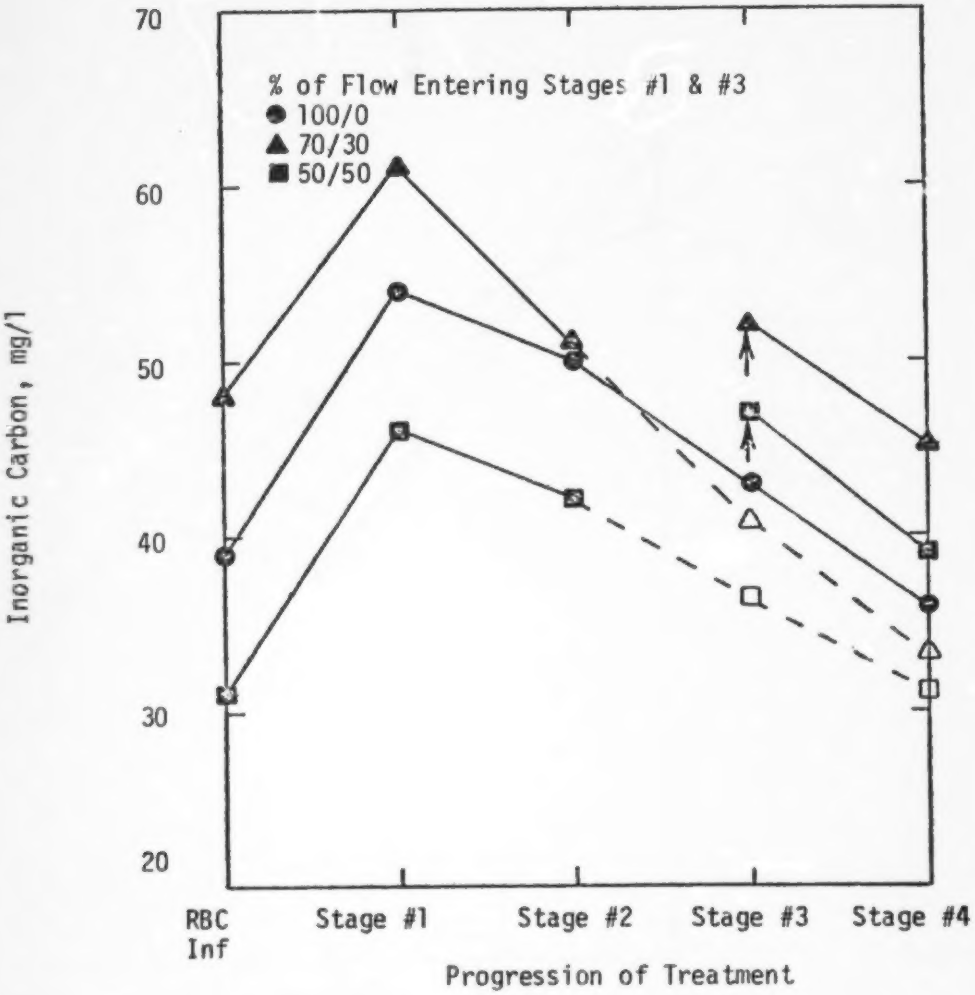

FIGURE 10. The Concentrations of Inorganic Carbon are Shown in the RBC Influent and Each Stage. The Symbols represent the Percent of Flow Entering Stages #1 and #3. Open Symbols are the Expected Values if Additional Wastewater had not been Applied to Stage #3.

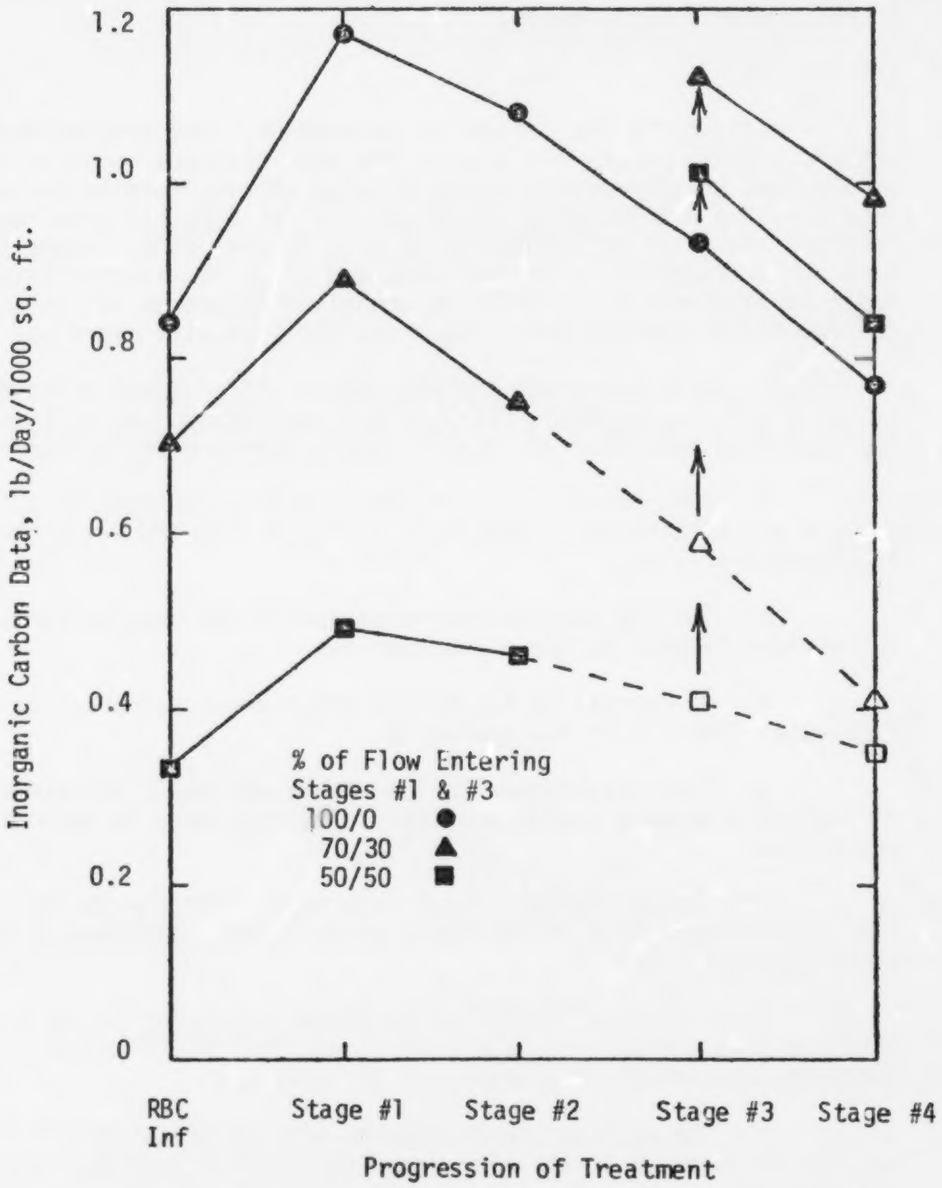

FIGURE 11. Inorganic Carbon Levels are Expressed as a Function of the RBC Surface Area and Time, Indicating Relative Rates of Production and Removal. The Symbols Represent the Percent of Flow Entering Stages #1 and #3. Open Symbols are the Expected Values if Additional Wastewater Had Not Been Applied to Stage #3.

The additional organic carbon applied to the third RBC stage was then converted to inorganic carbon increasing the observed rate of pH depression in the latter stages. From this data it can be concluded that flow splitting did allow for more efficient use of the RBC media surface area with respect to wastewater recarbonation.

CONCLUSIONS

In addition to the removal of phosphorus, lime pretreatment reduced the organic loading on the RBC process and provided recarbonation by microbial populations which produced carbon dioxide thereby forming carbonate alkalinity necessary for nitrification after the initial BOD_5 has been removed. The resultant pH after recarbonation is also in the optimal range for nitrification. Nitrification is further enhanced since the availability of the RBC media surface area for nitrifying organisms increased as the surface area requirements for heterotrophic organisms declines with decreased BOD_5 loading.

Under conditions where the wastewater influent was split between RBC stages 1 and 3, comparable rates of nutrient removal as well as overall recarbonation was observed. Other conclusions are as follows:

1. Phosphorus levels in the secondary effluent of 2.0 mg/l (as P) or less are attainable by combining low-level lime addition with biological treatment processes.

2. The RBC process can recarbonate a wastewater stream without deleterious effects to the microorganisms.

3. pH depression by the RBC process was decreased by increased flow rates from 2.0 to 4.0 gpd/sq ft.

4. Lime pretreatment allows for a decreased organic load on the biological treatment system and therefore a decrease in secondary sludge production.

5. The decreased organic load applied to the RBC as a result of lime pretreatment reduced RBC media surface area requirements for heterotrophic populations.

6. Biological oxidation of organics present in the high pH primary effluent produced an excellent environment for nitrification (low BOD, adequate alkalinity, pH between 7.0 and 8.0).

7. The application of high pH wastewater to the third RBC stage was not detrimental to the microorganisms or to the overall removal of nutrients.

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PREDICTION OF RBC PLANT PERFORMANCE FOR MUNICIPAL WASTEWATER TREATMENT

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INTRODUCTION

Fixed-film biological reactors have been a significant means to achieve secondary treatment of wastewaters discharged from both industrial and municipal sources. In the past, the most common was the trickling filter process. It was used primarily because it provided simplicity of operation, exhibited ability to handle wide variations of flow, and was proven a viable means of treatment (1).

Recently, due to more stringent regulations placing limitations on discharges, there has been an interest in another fixed-film wastewater treatment process called the rotating biological contactor (RBC) system. This system, relatively new to the United States, employed many of the advantages of the trickling filter process, along with the treatment efficiencies of the more advanced activated sludge system.

Generally, the RBC system consists of a series of large diameter disks constructed of plastic media, mounted on a horizontal shaft, set atop a concrete trough or tank. The disks on the shaft, positioned parallel or perpendicular to the flow, rotated slowly with approximately 40% of their surface area submerged in the wastewater. The organisms attach, grow and proliferated upon the disk surface. Excess growth or biomass is subsequently sloughed off by the shearing action of the disks as they move through the wastewater. This rotating movement serves a second function in which it causes a continual circulation of the wastewater and a dispersion of the excess biomass. Consequently, as the disks rotate a wastewater film is lifted out of the tank and slowly percolates down the attached biomass. Removal of BOD and nitrogen compounds takes place when sludge microorganisms consume both oxygen and substrate in the wastewater (2,3).

The disk media is usually arranged in a series of stages. In common practice, the RBC system varies from 4 to 6 stages dependent upon the design and expected treatment efficiencies. Past experience indicates that organic matter is often removed in the first stages of the process, with nitrogen removal in the subsequent stages. Therefore, it is known that the RBC system can be used to obtain significant carbonaceous as well as nitrogenous removals by using many combinations and/or arrangements of multiple RBC units, together or in conjunction with other processes. It lends itself quite well to the upgrading of the existing facilities (retrofitting) because of its modular construction, low head loss through the system, and required shallow excavation depth (4,5).

Additional advantages of the RBC system are economics, flexibility, simple operation and maintenance, suitability for step and stage construction, resistance to organic and hydraulic loads, low process control requirements, predictable performance, low sludge production, rapid settling and readily dewatered sludge, no requirement of aeration equipment, efficiency of oxygen transfer, less susceptability than activated sludge to washout and upset, no short circuiting, and easier acclimation to different wastes. However, the major disadvantages of the system are: (a) susceptability to vandalism and harsh weather environment unless covered, (b) relatively new technology and present employment for treatment of small wastewater flow, (c) possible odor problems (d) shaft and drive failure, and (e) oil leakage and required weekly grease (6,7).

To the best of our knowledge, the important controlling factors of the RBC system are considered to be the influent substrate concentration, surface hydraulic load, rotational disk speed, effective disk surface area, submerged disk depth, liquid retention time, wastewater temperature, and stage number (8).

Therefore, the prediction of the RBC wastewater treatment system is difficult because of the complexity of the process operating parameters.

The purpose of this study is four-fold: first, to correlate the efficiency of substrate removal to the process controlling variables; second, to develop both BOD and nitrification models for predicting the RBC plant performance; third, to compare the presently developed models with the others proposed earlier by Weng and Molof; and last, to verify the resulting models by using the data obtained from the operation of several full-scale RBC pilot plants.

MODEL FORMULATION

Mathematical models are usually derived and developed from either a theoretical basis or through a statistical analysis of experimental data. In the past, theoretical models for the descriptions of the relationship between substrate concentration and cell reproduction in the suspended-growth activated sludge system were commonly derived through the Monod equation. However, this equation cannot be directly applied to the RBC system because of the following two consequences. First, the varying quantity of biomass growth and sloughing in each stage makes the quantification of total biomass difficult and impracticable. Second, the biological utilization of substrate in the RBC system is attributed to the both attached- and suspended-growths. Therefore, the kinetic constants obtained from the Monod equation would be the overall rate constants, instead of those generally expressed for the attached growth system only (8,9).

The theory of film diffusion is another theoretical approach which has been used to predict the performance of the RBC system. This approach to model development is also unfeasible due to its complexity and low applicability; too many parameters are involved which may be difficult or impossible to obtain accurately on a routine basis (10). As a result of the impracticalities involved in using biokinetics and film diffusion theory to form a model of the RBC system, the multiple regression analysis technique was chosen.

Joost first suggested a model for predicting RBC plant performance as indicated in Eq. 1. However, his model did not have the support of either laboratory or field studies (11).

$$\%\ \text{Substrate Removal} = K\, L_o^{a}\, T^{b}\, S^{c}\, R^{d}\, N \text{ ----(1)}$$

in which

K = treatability constant of wastewater

L_o = influent substrate concentration of wastewater, in this case, either influent soluble BOD or ammonia nitrogen concentration

T = wastewater temperature

S = reactor residence time

R = physical configuration constant dependent upon disk diameter, spacing, thickness, and submergence

N = stage number

and a, b, c, and d are the partial regression coefficients.

In 1974, Weng and Molof proposed a model similar to that of Joost, however different in that the variables of flow rate (Q), disk surface area (A), disk rotational speed (B), submerged depth of disk (D) were included. The model was described as (12):

$$\% \text{ substrate remaining, } F = K L_o^a T^b S^c A^e B^f D^g Q^h \text{----------(2)}$$

In Eq.2, the L_o, T, S, K, a, b, and c are identical to those defined previously and the e,f,g, and h are new regression coefficients.

On the basis of Eqs.1 and 2, a general model employed for the prediction of the RBC system performance can be written as:

$$F = \xi L_o^a T^b S^c A^e B^f D^g Q^h \qquad \text{-----------------(3)}$$

in which $\xi = (K/N)R^d$. The value of ξ depends upon not only the nature of the wastewater to be treated, also the physical configuration of the RBS system including stage number, submerged disk depth, disk size and spacing.

Eq. 3 can be converted into a linear form by taking the logarithm of both sides of the equal sign.

$$\begin{aligned} \text{Log } F = {} & \text{Log } \xi + a \text{ Log } L_o + b \text{ Log } T + c \text{ Log } S \\ & + e \text{ Log } A + f \text{ Log } B + g \text{ Log } D + h \text{ Log } Q \end{aligned} \qquad \text{----------------------(4)}$$

or

$$\begin{aligned} \text{Log } F = {} & \P + a \text{ Log } L_o + b \text{ Log } T + c \text{ Log } S \\ & + e \text{ Log } A + f \text{ Log } B + g \text{ Log } D + h \text{ Log } Q \end{aligned} \qquad \text{----------------------(5)}$$

in which $\P = \text{Log } \xi$.

The investigative technique included collection of data from the literature and manipulation of the data by multiple regression analysis, all directed toward a model development. The computations of the data were performed on an IBM System 34 computer employing an IBM statistical analysis program. The program output provided a multiple correlation coefficient, partial regression coefficients for each independent variable, a constant term ¶, the number of degree of freedom, and a t-test.

MODEL DERIVATION

Model I - This model was developed for the prediction of carbonaceous BOD removal in RBC systems. The first study used for model development was conducted by Autotrol Corporation under field conditions with municipal wastewater at the Pewaukee Pollution Control Plant, Pewaukee, Wisconsin. The plant specifications are shown in Table 1.

Table 1. RBC Plant Specifications
For Pewaukee Study (13)

Disk Diameter	5.74 ft.
Disk Speed	2-5 rpm.
Number of Stages	2
Number of Disks	
Stage 1	45
Stage 2	46
Total Effective Disk Surface Area	4,600 ft.^{2}
Submerged Volume of Disks	420 gallons
Total Tank Volume	700 gallons

Seventy data sets that included influent BOD (L_o), flow rate (Q), disk rotational speed (B), liquid retention time(S), effective disk surface area (A), and wastewater temperature (T), were employed for multiple regression analysis. However, the submerged disk depth (D) was not considered as a process variable in this study because it was maintained constant throughout the entire experiment. Table II. shows the results of the multiple regression analysis.

Table II. Regression Analysis of RBC Variables For Pilot Plant Data Obtained From Pewaukee Study (Using Eq.5)

FACTOR	VARIABLE					
	L_o	Q	B	S	A	T
Regression Coefficient	-0.6657	0.5844	-0.0589	-	-8.1573	-0.2587
t-test	-2.7837	7.1970	-0.0344	-	-1.0252	-2.3221

$\P = 1.9325$, $R_c = 0.8847$, DF = 18.0, $t_{0.975} = 2.101$

The regression coefficient for Log B did not contribute to regression as shown by the "t-test" in Table II. In addition, the correlation matrix for the data shown in Table II reveals that the liquid retention time (S) is co-linear with flow (Q) as would be expected from the method of testing the RBC system. Thus, the liquid retention time was eliminated from the list of the process variables. According to the data shown in Table II, the resulting model is:

$$F = \frac{1.9325 \; Q^{0.5844}}{L_o^{0.6657} \; B^{0.0589} \; A^{8.1573} \; T^{0.2587}} \text{-------}(6)$$

It is known from Eq.6 that the major controlling factors in the RBC system consist of the wastewater flow (Q), influent BOD concentration (L_o), disk surface area (A), and wastewater temperature (T). By deleting B, S, and D from the group variables and using q as hydraulic loading for replacing Q and A (i.e., Q/A), Eq.5 becomes

$$\text{Log } F = \P + a \text{ Log } L_o + b \text{ Log } T + r \text{ Log } q \text{ ---------}(7)$$

In Eq.7, the r is a new partial regression coefficient.

The multiple regression analysis was run again by using Eq.7 and the results are summarized in Table III. The expression derived from the data in Table III and Eq.7 by regression analysis is shown in Eq.8.

Table III. Regression Analysis of RBC Variables For Pilot Plant Data Obtained From Pewaukee Study (Using Eq.7)

FACTOR	VARIABLE		
	L_o	q	T
Regression Coefficient	-0.6837	0.5579	-0.2477
t-test	3.0449	7.4898	-2.5690

$\P = 7.55$, $R_c = 0.8904$, $DF = 18.0$, $t_{0.975} = 2.101$

$$F = \frac{7.55\, q^{0.5579}}{L_o^{0.6837}\, T^{0.2477}} \quad \text{at } N = 2 \qquad \text{------------------} (8)$$

Eq.8 has an exponent of 0.6837 for the variable L_o, which indicates that L_o is the most significant variable. Also, the value of ¶ in Eq.8 is equal to 7.55. It is important to emphasize the fact again that the ¶ is a function of stage number, disk and wastewater properties. So, an attempt was made to determine the effect of stage number N on ¶ value in Eq.7.

The data resulted from the operation of five RBC plants as specified in Table IV were used to investigate the relationship between ¶ and N. The value of ¶ was calculated from Eq.7 at the exponent a = 0.6837, b = -0.2477, and r = 0.5579. The calculated ¶ was then plotted against the stage number N as shown in Figure 1.

Table IV. RBC Plant Specifications From References 7, 9, 10, 13, and 14

	Reference Number				
Parameters	7	9	10	13	14
Disk Diameter, ft	6.56	12.0	3.0	12.0	4.0
Disk Speed, rpm	2.90	--	--	--	6-10
Number of stages	4.0	4.0	10.0	6.0	4.0
Number of disks	--	--	480	--	--
Total surface area, ft^2	7,900	196,500	6,200	516,000	3,770
Tank volume, ft^3	900	18,800	2,400	62,500	500

It can be seen in Figure 1 that the ¶ decreases with an increase in the number of RBC stages employed. This result has clearly demonstrated that the biodegradability of wastewater is gradually decreased in each succeeding RBC stage. The same finding was also reported by Antonie, Sack, and Pretorias (8, 22, 25).

A first-order relationship between ¶ and N was found in Figure 1 and the descriptive equation is

$$¶ = 14.2 \text{ Exp.}^{-0.32N} \quad \text{------------(9)}$$

The data points follow Eq. 9 well when the RBC system has six stages or less. However, the calculated ¶ falls below the observed values after the stage number exceeds six stages.

Substituting Eqs.8 and 9 into Eq.7, the general BOD model for RBC systems is obtained

$$F = \frac{14.2 \; q^{0.5579}}{\text{exp.}^{0.32N} \; L_o^{0.6837} \; T^{0.2477}} \quad \text{----------(10)}$$

It is known from Eq.10 that the BOD removal capacity of the RBC system depends mainly upon the following important parameters: hydraulic loading, initial soluble BOD concentration, wastewater temperature, and RBC stage number. The final relationship among these factors is that the soluble BOD remaining F is inversely proportional to the influent soluble BOD concentration L_o, wastewater temperature T and RBC stage number N, but is directly proportional to the hydraulic loading rate q.

Model II - This model was developed for the prediction of nitrogen removal in RBC systems. Previous work conducted by Wild, Sawyer, and McMahon showed that the rate of biological nitrification was significantly affected by the organic content of the wastewater (17). For this reason, the RBC nitrification model was determined by the selection and analysis of data in which the associated influent soluble BOD content in the wastewater was either greater than 80 mg/l or less than 50 mg/l. This was intended to limit the application of the presently developed nitrification models to either a RBC secondary treatment system or a RBC tertiary treatment system.

Wastewater treatment study for nitrification of primary effluent conducted by Autotrol Corporation at the Pewaukee Water Pollution Control Plant, Pewaukee, Wisconsin was used to develop the nitrification model when the RBC unit served as a secondary treatment system. The plant specifications of the Pewaukee RBC facility were alreay reported in Table 1.

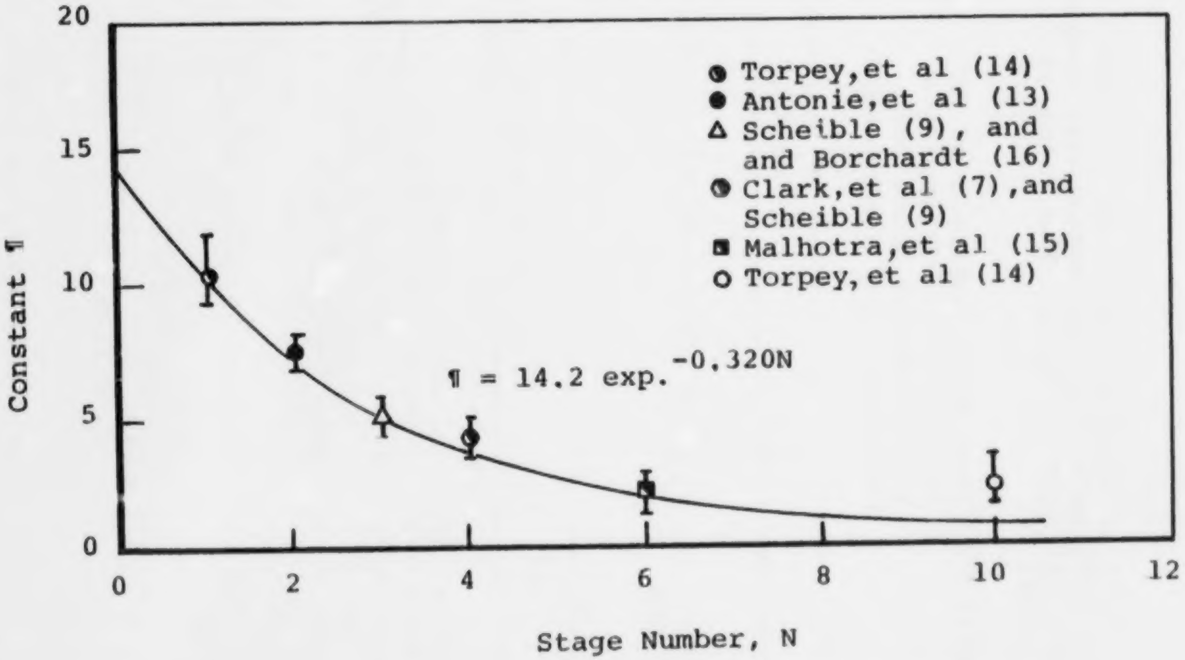


Figure 1. ¶-N Relationship For BOD Model In A RBC Wastewater Treament System

The total number of data sets used for the study of this model was 101 and the multiple regression analysis of the data was based on Eq.7. At the stage number N = 4, the ammonia nitrogen remaining in the RBC plant effluent can be stated by Eq.11 in accordance with the data shown in Table V.

$$F = \frac{4.336\ q^{0.1692}}{L_o^{0.2395}\ T^{0.6229}} \quad \text{-------(11)}$$

The greatest value of an exponent in Eq.11 is 0.6229 for the variable T. This indicates that the wastewater temperature is the most important variable in Eq.11. The value of ¶ was found to be equal to 4.336.

Table V. Regression Analysis of RBC Variables In Developing A Nitrification Model From Pewaukee Study (Using Eq.7)

FACTOR	VARIABLE		
	L_o	q	T
Regression Coefficient	-0.2395	0.1692	-0.6229
t-test	-0.6706	1.1430	-3.3390

$¶ = 4.3366$, $R_c = 0.9518$, $DF = 12.0$, $t_{0.975} = 2.447$

The experimental data collected by Wilcox et al at Belmont RBC pilot plant were used to study the nitrification model for a RBC tertiary treatment system. The plant specifications of the Belmont RBC facility are indicated in Table VI. The multiple regression analysis of the data resulted from the above study were run by using Eq.7 and the results are shown in Table VII.

Based on the results reported in Table VII, Eq.7 can be written at N = 4 as

$$F = \frac{1.552 \times 10^{-4}\ q^{0.6803}\ L_o^{2.6452}}{T^{0.2170}} \quad \text{-----(12)}$$

The largest value for an exponent in Eq.12 is 2.6452 for the variable L_o. This means that the influent ammonia nitrogen concentration is the most important controlling variable in this model. Also, the value of ¶ equals to 1.552×10^{-4}, according to Eq.12.

There is a remarkable difference between Eq.11 and Eq.12 and this variation is primarily caused by the external restriction of influent soluble BOD in the wastewater. Additionally, it can be seen in Eqs.11 and 12 that the effect of hydraulic loading q, influent ammonia nitrogen concentration L_o, and wastewater temperature T on the effluent ammonia nitrogen remaining F is different. In Eq.11, the F is inversely proportional to L_o and T, but in Eq. 12 the F is inversely proportional to T only. Nevertheless, it was found from this study that the relationship between F and q was always maintained at the same, that is an increase in q results in an increase in F.

Table VI RBC Plant Specifications For Belmont Study (18)

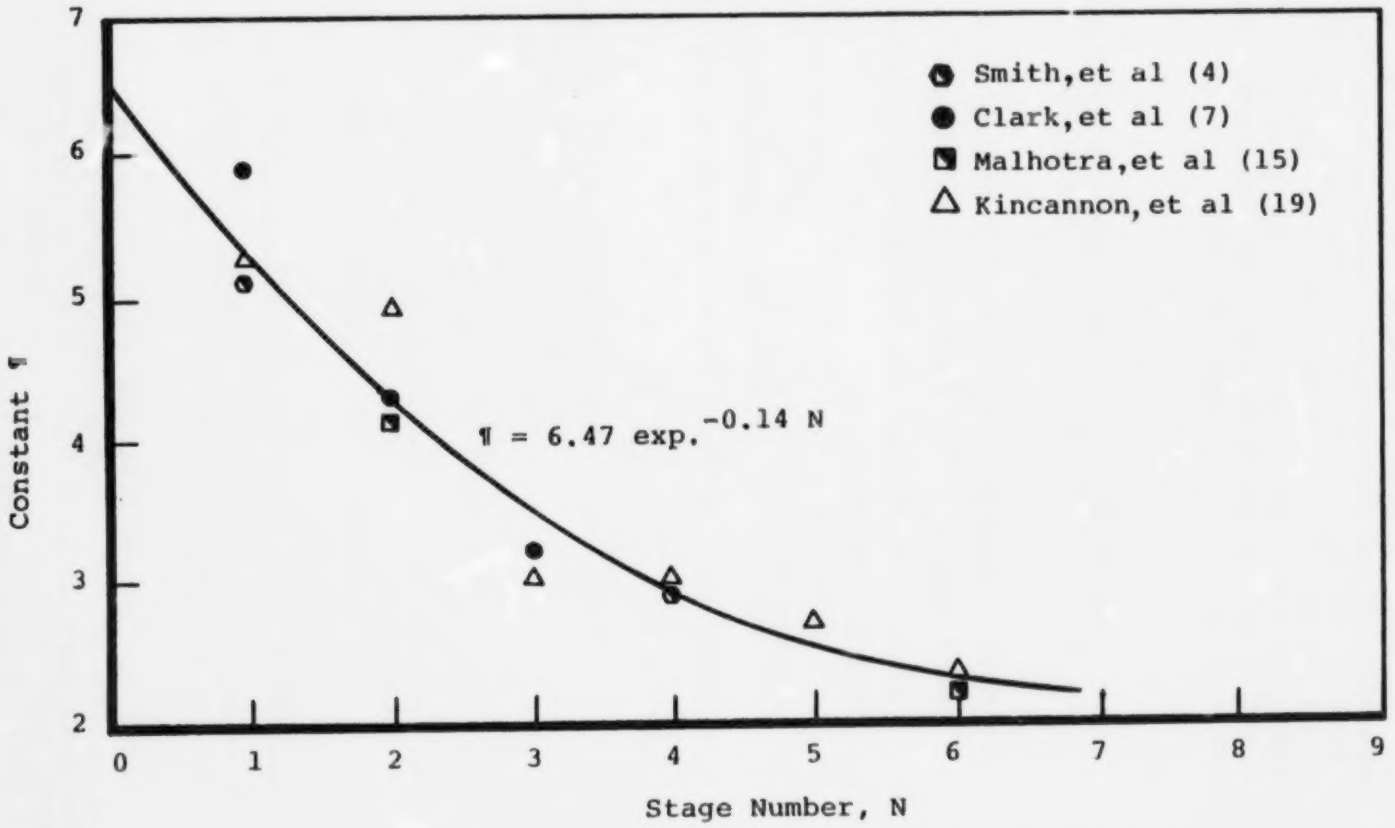
Disk Diameter	12.0 ft
Disk Speed	0.8-1.8 rpm
Number of Stages	4
Total Disk Surface Area	22,628 ft^2
Submerged Tank Volume	4,020 gallons
Total Tank Volume	8,758 gallons

Table VII. Regression Analysis of RBC Variables In Developing A Nitrification Model From Belmont Study (Using Eq.7)

FACTOR	VARIABLE		
	q	L_o	T
Regression Coefficient	2.6452	0.6803	-0.2170
t-test	6.5020	2.2720	-0.3096

¶ = 1.552×10^{-4}, $R_c = 0.9755$, DF = 6.0, $t_{0.975} = 2.571$

In order to investigate the ¶-N relationship for the process of nitrification in RBC systems, eight literature sources were used. The data were entered Eq.7 either at the exponent a= 0.2395, b= -0.6229, and r= -0.1692 or at the exponent a= 2.6452, b= -0.2170, and r= 0.6803, depending upon the soluble BOD concentration in the



Fifure 2. ¶-N Relationship In A RBC Secondary Wastewater Treatment System, Influent Soluble BOD Concentration>80 mg/l

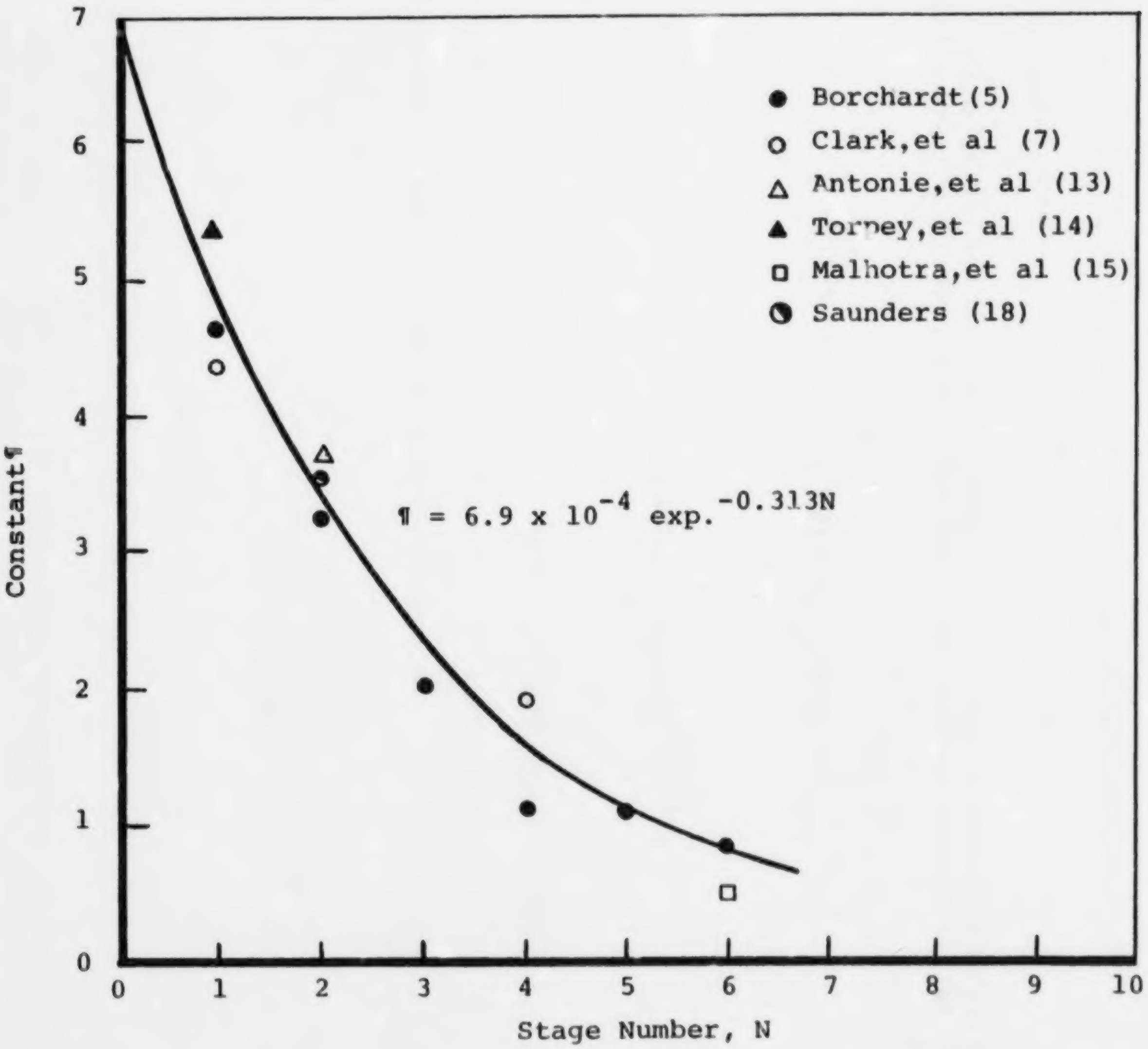


Figure 3. ¶-N Relationship In A RBC Tertiary Wastewater Treatment System, Influent Soluble BOD Concentration< 50 mg/l

influent wastewater. The calculated ¶ against the stage number N was then plotted in Figures 2 and 3.

The results as shown in both Figures 2 and 3 indicate that the ¶ decreases with increasing the number of RBC stages employed. Based on the data presented in Figures 2 and 3, the ¶-N relationship can be described by two first-order equations as follows:

$$\P = 6.47\ \mathrm{Exp.}^{-0.140N} \text{ ------------------ } (13)$$

and

$$\P = 6.90\ \mathrm{Exp.}^{-0.313N} \text{ ------------------ } (14)$$

Substituting Eqs.11 and 13 into Eq.7 or Eqs.12 and 14 into Eq.7, the RBC nitrification models are obtained. The descriptive equations are:

$$F = \frac{6.47\ q^{0.1692}}{\mathrm{exp.}^{0.140N}\ L_o^{0.2395}\ T^{0.6229}} \text{ ------------ } (15)$$

$$F = \frac{6.90\mathrm{x}10^{-4}\ q^{0.6803}\ L_o^{2.6452}}{\mathrm{exp.}^{0.313N}\ T^{0.2170}} \text{ ------------ } (16)$$

MODEL COMPARISON AND VERIFICATION

Weng et al have earlier proposed two mathematical models similar to Eqs.10,15 and 16 (8 , 12). Their models for the prediction of BOD and nitrogen removals in RBC systems, respectively, are:

$$F = \frac{2.4465\ L_o^{0.2815}\ Q^{0.4533}}{S^{0.3988}\ A^{0.7004}\ T^{1.2423}} \text{ -------------- } (17)$$

$$F = \frac{0.0545\ L_o^{0.6440}\ Q^{0.1440}}{S^{0.5300}\ A^{1.2760}} \text{ -------------- } (18)$$

The main differences among Eqs.10, 15, 16, 17, and 18 are: (a) the number of RBC stages and the wastewater temperature are not included in Weng's model but the operating parameter of disk rotational speed is considered, (b) the flow and disk surface area are synonomous with the hydraulic loading in Eqs. 10, 15 and 16 as q=Q/A, and (c) the partial regression coefficient values for the Weng's model and the models obtained from the present study are different.

The stage number and the wastewater temperature are the significant variables which should be considered in the modeling of RBC systems. All models proposed by Weng et al did not include temperature variable because they were developed from data at a constant temperature and therefore regression analysis reported it as not significant. Additionally, exclusion of stage number as an independent variable will not allow usage of the model for comprehensive design purposes. The rotating disk speed is included as a model parameter in Eqs.15 and 16 and although it is important for RBC operation it varies only slightly between 1-4 rpm, as it is presently designed into the process of the manufacturer.

The experimental data reported by Hao and Hendrick (21), Sack, et al (22), and Antonio (6) were taken to test Eqs.10 and 17, and the results are presented in Figure 4. It is apparent from Figure 4 that the BOD model developed from this study yields better agreement than that resulted from the Weng's model.

In addition, data from the studies of Wu (23), Wilcox (18), and Marsh (24) were used to examine the nitrification models as proposed in Eqs.15, 16, and 18. The above studies were conducted by using the full-scale RBC plants having a disk size > 10 ft. The verifications of Eqs.15 and 16 are illustrated in Figures 5 and 6 respectively. However, the results obtained from the Weng's nitrification model were not shown in Figures 5 and 6 because the predicted values were greatly different from the observed data.

CONCLUSIONS

The results of this study show that:

If the survival and growth of sludge organisms in the rotating biological contactor wastewater treatment systems is independent of pH and alkalinity, the efficiencies of BOD and ammonia nitrogen removal can be easily predicted by simply knowing of stage numbers, hydraulic loading, influent substrate concentration, and wastewater temperature. Both BOD and nitrification models presently developed yield good agreement between calculated results and observed data.

According to the present study, the BOD model is given as:

$$F = \frac{14.2\, q^{0.5579}}{\exp.^{0.32N}\, L_o^{0.6837}\, T^{0.2477}}$$

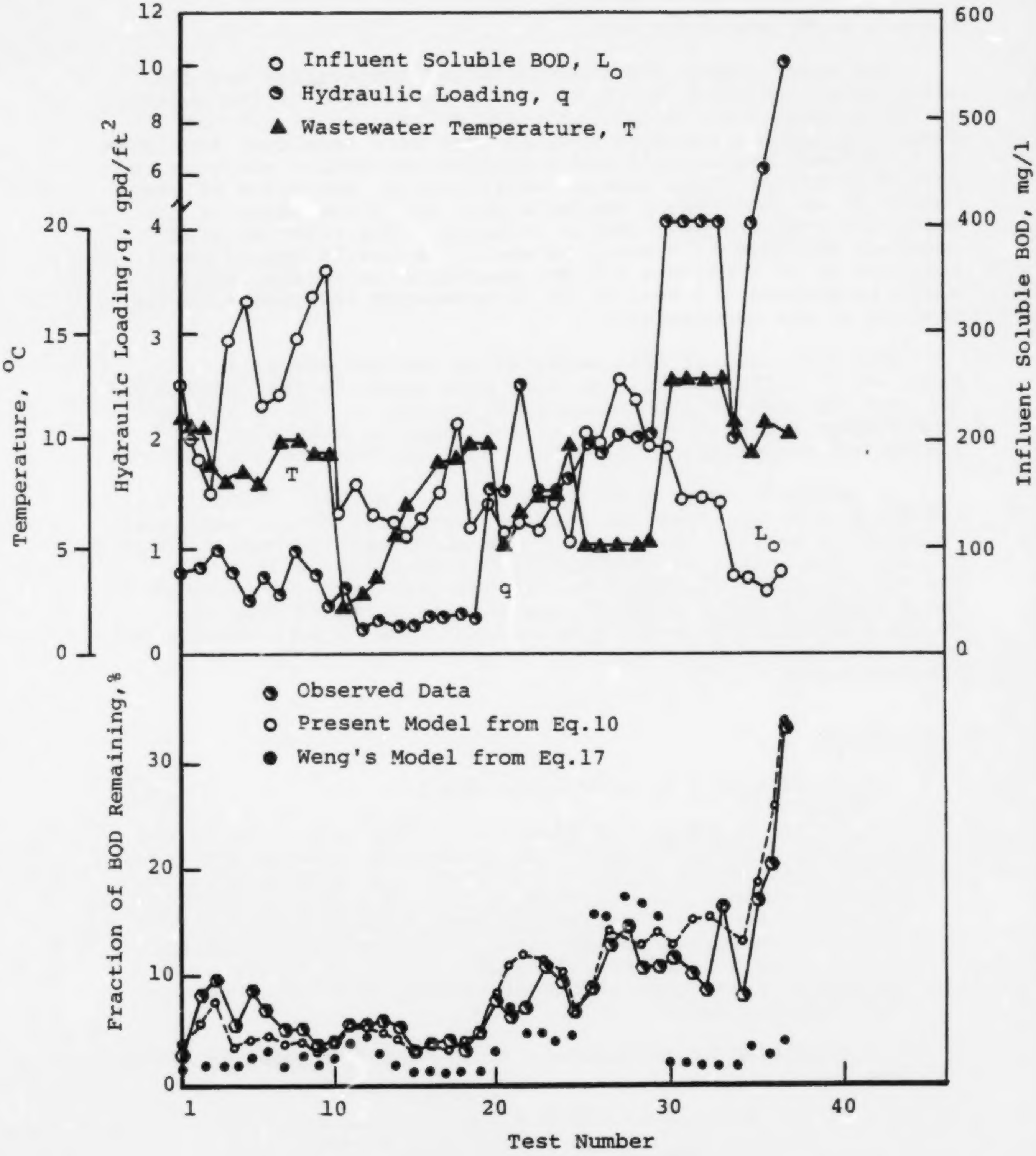


Figure 4. Field Observations and Model Predcitions From Eqs. 10 and 17

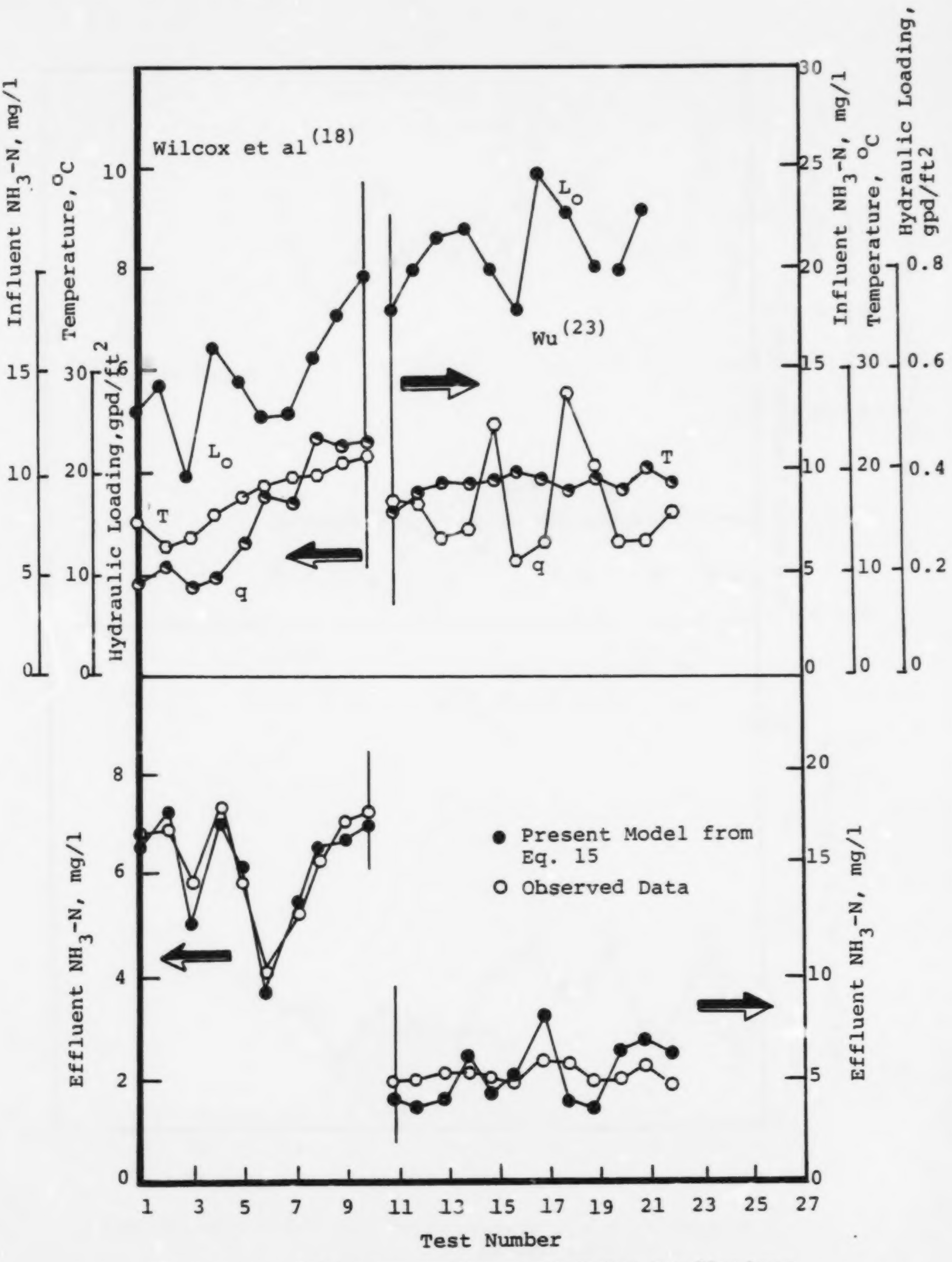


Figure 5. Field Observations and Model Predictions From Eq. 15

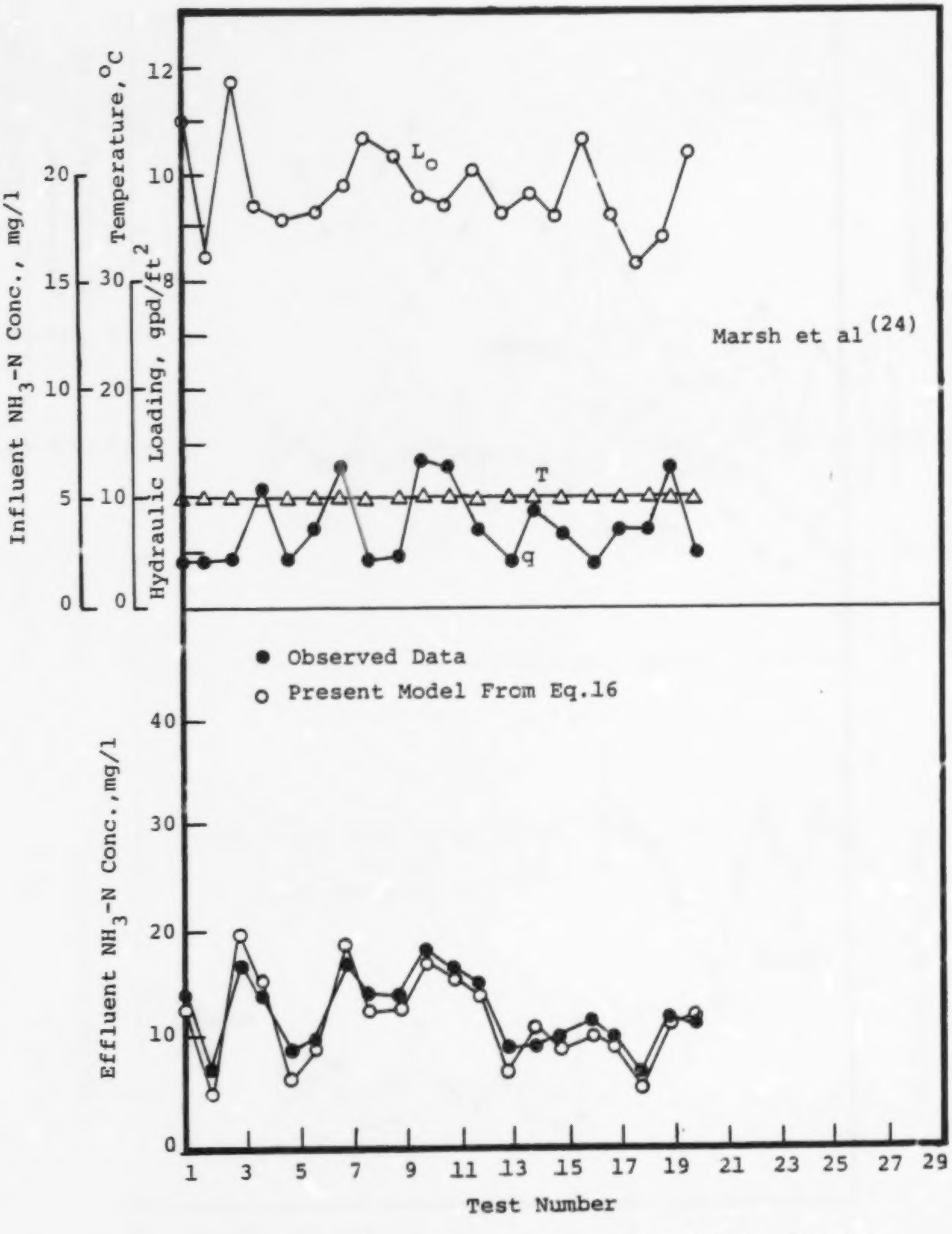


Figure 6. Field Observations and Model Predictions From Eq. 16

When the RBC plant is operated to serve as a secondary treatment system, the nitrification model is

$$F = \frac{6.47 \ q^{0.1692}}{\exp.^{0.14N} \ T^{0.6229} \ L_o^{\ 0.2395}}$$

In addition, the nitrification model for RBC tertiary wastewater treatment system is

$$F = \frac{6.90 \times 10^{-4} \ q^{0.6803} \ L_o^{\ 2.6452}}{\exp.^{0.313N} \ T^{0.2170}}$$

The significant aspect of this study is that it demonstrates the feasibility of using the RBC to control the removal of carbonaceous BOD and ammonia nitrogen from municipal wastewater and to optimize the utilization of disk surface area under a wide range of operating conditions. These models presently obtained are relatively simple and can be used by the environmental engineer to design RBC systems without knowledge of kinetic information.

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Notations

The following symbols are used in this paper:

A = total effective disk surface area, ft^2

B = disk rotational speed, rpm

D = submerged disk depth,%

DF = degree of freedom

F = fraction of influent soluble BOD or ammonia nitrogen remaining in the effluent

K = treatability constant of waste material

L_o = influent concentration of soluble BOD or ammonia nitrogen, mg/l

N = number of stages

Q = flow rate,gpd

q = hydraulic loading, gpd/ft^2

R = physical configuration constant, depends upon disk diameter, spacing thickness, submergence, etc.

R_c = multiple correlation coefficient

S = reactor residence time, hours

T = temperature, ^oC

$\xi = (K/N) \times R^d$

¶ = Log ξ or Log $(K/N)\ R^d$

a,b,c,d,e,f,g,h,and r
= partial regression coefficients

RBC DESIGN CONSIDERATIONS FOR INDUSTRIAL WASTEWATERS

by

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INTRODUCTION

The decade of the 1970's can be described as the "decade of change" with respect to industrial wastewater management. Major changes in environmental goals were reflected by the introduction and current common usage in our working vocabulary of terms such as ammonia nitrogen, nitrate nitrogen, BOD_{28}, phosphorus, and a host of priority pollutants as well as what once would have been considered unbelievably low requirements for the more familiar BOD_5 and suspended solids parameters for describing effluent design criteria. As we leave the decade, the concept of COD as an "indicator pollutant" has recently been proposed as a regulatory parameter (1). If enacted, this new regulatory concept may have more impact on some industrial wastewater treatment systems during the new decade than the impact of the new regulatory requirements enacted during the previous decade.

These new regulatory concepts have been accompanied by significant changes in both cost philosophy and treatment process understanding. During this decade, our approach to industrial waste treatment has changed from "It can't be done (because of the cost)!" to a philosophy of "Whatever it costs, it will be done!" Research during the decade has provided new information and concepts that in turn have yielded better understanding about the abilities and limitations of treatment processes.

As a result of innovative thinking and research, the decade of the 1970's saw the development and introduction of several new processes for wastewater treatment. The development and widespread implementation of rotating biological contactors probably represents the most significant new treatment process introduced during the decade. However, several RBC failures have created concern about process reliability among engineers trying to meet the challenge of designing cost and energy effective industrial wastewater treatment systems. The primary purpose of this paper is to describe the major causes of process failure so that engineers can take full advantage of the cost effective potential of RBC treatment without the fear of process failure.

In the early 1970's RBC technology was a developing art with only a few experimental units available to provide design information for domestic wastewater treatment situations. Industrial utilization of this innovative and energy saving process was based on trial and terror practices which resulted in a few notable failures along the way. In the early part of the decade, sales personnel claimed that RBC's could treat virtually any biodegradeable wastewater, under any conditions, to any desired degree of treatment. All that was required was sufficient media! Claims were also made that the process (1) had a very high resistance to toxic slugs and shock loads; (2) could be built at lower costs than alternative treatment systems; (3) could be operated with substantial power savings compared to alternative treatment systems; (4) produced minimal solids yield; and (5) required minimal operator attention. For design engineers accustomed to hydraulic detention times of six to twenty-four hours the two hour RBC hydraulic detention time seemed unbelievable, although it proved to be true for many situations. Unfortunately, RBC's have not proven to be quite the panacea promised, especially for industrial wastewaters. Even though they have been applied to or pilot-tested with a large variety of industrial wastewaters with a great deal of success, failures with some industrial wastes have occurred and will probably continue to occur in the future. RBC's have distinct fundamental limitations that can lead to process failures for some wastewaters. However, these limitations are rarely described in the literature. The primary purpose of this paper is to categorize and describe some of these failure modes and to suggest preventive evaluation procedures for the consideration of designers working with industrial wastewaters.

FAILURE MODES

RBC failure can be defined as any situation where the process does not meet effluent goals or does so in an objectionable manner. Situations such as process inability to meet effluent BOD and/or ammonia standards, or the production of solids that won't settle or cannot be readily separated from the carrier stream, or the production of objectionable odors are examples of process failure modes. Obviously, media separation, shaft, bearing and mechanical drive train problems are also previous causes of process failure. Although designers should be able to learn from previous RBC failures, there is a paucity of available descriptive information for the following reasons:

1. Design engineers and consulting firms avoid describing their failures for fear of degrading their public images.
2. Equipment manufacturers won't voluntarily describe previous mechanical failures of their equipment or process failures for fear of interference with future sales. Previous failures are indirectly alluded to when manufacturers introduce their "new improved whatchamacallit" or "thingamajig variation" in their sales literature.
3. Facility owners are understandably unhappy when their wastewater treatment system fails. Owners try to avoid having their public image associated with or recognized in conjunction with failures. Frequently,

based on the advice of their attorneys, owners refuse to discuss or permit interested outside parties to visit and inspect these failures for fear of influencing pending litigation. Smaller industries are deeply concerned about their capital loss and their relationship with regulatory agencies. Larger organizations, having more capital resources, salvage what they can, bury their mistakes and go on to a new treatment system.

Regardless of the reason for silence, it is human nature to openly describe successes and forget failures. Several owners and engineers contacted by the author of this paper were reticent to discuss their process failures because of "corporate policy." Hence, failure data are not generally available for inspection and examination. However, failure descriptions are passed around by the design profession by word of mouth or by pirated copies of clandestine letters, memoranda and draft reports intended for "internal use."

There are three general classes of in-plant problems: (1) process, (2) equipment and (3) operational failures. Process failures result from inadequate or insufficient pilot testing and poor planning by the design engineer. Equipment failures are less likely in the future, as the major RBC manufacturers have modified their products to eliminate major equipment design problems. Operational problems occur due to inadequate maintenance, or more frequently when the loading conditions are changes without regard to the effects of these changes on the wastewater treatment system. These three failure modes may be interactive. For example, significantly increased organic loading can result in abnormally heavy biomass attachment to the RBC discs which in turn may result in stressing equipment to levels far beyond those anticipated during design. Effluent quality will also probably decrease as a result of the increased loading. Thus, an operational change may bring about what could be incorrectly described as equipment and process failures.

PROCESS DESIGN CONSIDERATIONS

The following examples of RBC design factors are each related to at least one prototype RBC failure or potential failure. Industrial owners' representatives and design engineers, while willing to discuss their problems privately and "off the record," requested that data and/or identifying wastewater characteristics be held in complete confidentiality for the reasons previously cited. However, bench and pilot plant RBC data from the open literature will be used to illustrate the types of problems encountered. Most industrial RBC process design problems can be attributed to (1) unique characteristics of the industrial wastewater, (2) loading conditions, or (3) improper pilot plant operating conditions and poor data interpretation.

Wastewater Characteristics

RBC's are primarily effective for removing soluble wastewater constituents. Particulates entering the RBC unit will either pass through the unit without treatment or become enmeshed in the biomass attached to the discs. Because the hydraulic detention time in the RBC unit and its clarifier is so short, small influent organic particles escaping capture in the treatment train can exert high effluent oxygen demands when measured by standard BOD tests. Alternatively, if enough influent particles are trapped on the discs, (1) structural problems due to the extra weight can cause mechanical failures; (2) particle bridging between adjacent discs will prevent the distribution of substrate and more importantly the transfer of oxygen to the biomass, thus resulting in process failure; or (3) organic particles enmeshed within the biomass can create localized oxygen demands that exceed the oxygen transfer capability of the system and lead to anaerobic conditions. Thus, RBC systems to be designed for wastewaters containing significant quantities of particles should always be preceded with effective particle separation facilities.

Wastewater constituents should carefully be analyzed for a wide range of plant production schedules and conditions. While composite samples are useful for evaluating average design conditions, a careful plant analysis is required to account for peak organic concentrations and maximum flow conditions, along with nutrient balance assessments for these peak conditions. Because RBC systems are less forgiving than alternative suspended culture systems having longer hydraulic detention times (2, 3, 4). equalization should be strongly considered for systems subjected to frequent peak loading conditions in order to prevent undesirable transients in effluent quality.

Industrial wastewaters should be very carefully scrutinized for constituents that may form precipitates in the RBC units as a result of oxidation and/or pH changes. Recent bench scale RBC studies with tannery beamhouse wastewaters in the Sanitary Engineering laboratories at Syracuse University's Department of Civil Engineering serve to illustrate this type of potential problem. After several months of continuous operation, an inorganic precipitate probably consisting of a calcium sulfate-calcium carbonate complex was found throughout the biomass attached to the discs. Even though this one-quarter to three-eighths inch thick layer did not appear to interfere with soluble COD removal for these experiments, it is likely that the development of similar biomass-precipitate combinations could lead to unanticipated mechanical stresses, additional power requirements and possible bridging conditions in prototype units designed solely on a basis of areal substrate removal criteria. For situations of this type, either additional pre-treatment or an alternative treatment system should be recommended.

The nature of the organics in the wastewater can have a severe impact on the choice of treatment system to be employed. Readily biodegradeable soluble organic materials are ideal substrates for the RBC process. However, some industrial wastewaters present special problems that may not be recognized until construction is complete and several months of operation have transpired.

Figure 1 presents steady state operating data obtained with an eight-stage, bench scale RBC unit treating tannery beamhouse wastes following neutralization, sedimentation and anaerobic treatment. Each stage had about 4.4 square feet of disc area and the total unit had a surface area of 35.2 square feet and a nominal hydraulic detention time of 17.7 hours. As with many high strength industrial wastewaters, soluble COD is the parameter of choice due to better precision and repeatability than that obtained with the standard BOD_5 test. Conventional interpretation of these data would suggest that biological treatment was essentially complete by the end of the third stage and that the overall removal rates in these three stages were about 5.6 and 3.3 pounds of COD and BOD_5 per thousand square feet, respectively. The reactions appear to have been completed in about 6.6 hours and the overall removals are about 83 and 99 percent for COD and BOD_5 respectively. Excess oxygen and nutrients along with near neutral pH conditions in the succeeding stages tend to confirm the conclusion of a complete reaction. The effluent soluble COD, about 260 mg/l, for this situation would be termed refractory and ignored for design purposes. However, long term BOD tests, conducted with seed microorganisms from the RBC effluent yield the data shown in Figure 2. As a result of this additional information, it becomes obvious that the biological reactions did not go to completion in the time available in the RBC reactor Thus, effluents discharged from a system designed on data similar to those shown in Figure 1 could readily meet most effluent BOD_5 criteria but would also result in unexplained oxygen demands in receiving waters.

The most probable explanation for the apparent paradox presented by the data for this industrial wastewater is that readily biodegradeable wastewater constituents are removed in a few hours in the RBC unit but the complex soluble organic materials (probably proteins) remaining in the carrier stream require more time to be enzymatically broken down into biodegradeable units than is available within the RBC reactor. Other industrial wastewaters exhibit similar slow biological degradation characteristics and each wastewater should be evaluated as described above prior to process selection and design. Where short RBC hydraulic residence times are not adequate for complete treatment, alternative treatment methods, including hybrid RBC systems, should be evaluated.

Loading Rates

Overall RBC performance with industrial wastewaters is dependent on hydraulic and organic application rates as well as the nature of the wastewater. Process performance at the microbial level is usually controlled by oxygen and substrate transport into and through the biofilm and/or the transport of waste products away from reaction sites (5, 6). Organic loading rate in terms of mass applied per unit area per unit time seems to be the most appropriate design parameter for predicting effluent quality. Loading rates between about one-half and three pounds of soluble BOD_5 per 1000 square feet per day will generally produce an effluent containing less than 20 mg/1 of soluble BOD_5 along with readily settleable secondary solids. However, higher loading rates result in decreasing effluent quality and adverse operating conditions. Surprisingly, evidence is becoming available which indicates that very low loading rates (less than one-half pound $BOD_5/1000\ ft^2$-day) do not necessarily assure an outstanding effluent quality. This is probably due to lack of a substrate concentration sufficient to drive organic compounds into the biomass film. Conversely, at organic loading rates of about five pounds soluble BOD_5 per 1000 square feet per day, RBC reactors begin to behave as a combined fixed film-suspended culture system. Under these loading conditions effluent quality deteriorates in terms of both soluble BOD_5 and settleable suspended solids. Also, strong unpleasant odors are likely to be generated as rapidly growing suspended microbes deplete the available oxygen in the bulk carrier stream.

These adverse phenomena are illustrated by the following data obtained with an eight stage, bench scale RBC unit treating a soluble readily biodegradeable wastewater under ideal laboratory conditions. Figures 3 and 4 present data obtained under steady state operating conditions with disc rotational speed being the only operating parameter varied for this set of experiments. Complete data sets from these experiments have been reported elsewhere (5). Figure 3 indicates the soluble COD remaining in each stage and Figure 4 shows the stage dissolved oxygen profiles for the same loading conditions. Dispersed growth particles were found in each stage and in the effluents for all four experiments. The overall loading rate was about 5 lbs. of BOD_5 (7.6 lbs. of COD) per 1000 square feet per day. The units had insufficient oxygenation capacity to meet the biological oxygen demand at the lower two rotational speeds. This in turn encouraged biological sulfate reduction and generation of strong hydrogen sulfide odors from the unit. However, more than adequate dissolved oxygen was available for the higher two rotational speeds and is reflected in the better soluble COD removal for these conditions. Note that the first stage(s) is where the critical operating condition occurs. These data suggest that bypass or step feeding, supplemental air, or increased rotational speed provided by variable speed drive trains may be useful during peak loading conditions. Similarly, the removal of baffles between the first two (or more) stages for overloaded prototype plants may help to alleviate anaerobic conditions by increasing the oxygenation capacity of the first stage.

Pilot Plant Recommendations

Each industrial wastewater stream is unique. The experiences and information obtained from a plant at one location producing a particular product line, while providing useful guidelines, rarely can be transferred to another plant producing a similar product. Well designed RBC pilot plant programs can help engineers to design systems that will provide years of low cost, trouble free operation. Conversely, a poor or incomplete pre-design evaluation program can lead to years of continuous operational problems and frustrations. Good pilot testing programs are not inexpensive, but they are much less costly than system failure and subsequent remedial actions.

Both bench scale and prototype pilot plant testing are complementary and have preferred applications. Bench scale laboratory studies can be used to define the upper limits of treatability for a specific industrial wastewater stream. Critical conditions involving factors such as precipitate formation, toxic conditions, temperature effects, or the effects of mixing or segregating various in-plant wastestreams can be effectively evaluated with bench scale units under controlled conditions. Unfortunately, experimental lead time is usually critical in the design process and preliminary studies are frequently rushed to completion. The acclimation and replacement of biomass may take up to several months to develop, mature and produce meaningful results following changes in loading parameters. This is especially important where the effects of changed influent loading conditions can be anticipated to affect and interfere with nitrification in latter stages. Complete stage by stage data sets including dissolved oxygen, pH, alkalinity, nitrogen, solids and the concentrations of compounds of specific interest, in addition to stage BOD or COD data are highly desirable for defining the reactions and interactions taking place in the RBC unit. Daily scraping is necessary to prevent excessive wall growth from interfering with data analyses and interpretations. Effluent data should include an evaluation of solids settling characteristics as well as the parameters previously described.

Where a wide variety of process variables are to be evaluated over an extended range of values, multiple bench scale experimental units should be employed. While these recommendations are costly to implement compared to standard quick input-output experiments, they should prevent downstream process "surprises" that might otherwise be attributed to undefined "changes in the wastewater characteristics between pilot studies and construction."

Unfortunately, bench scale pilot plants cannot duplicate the hydraulic and oxygen transfer characteristics of full size RBC units (4, 5). Portable full size RBC units can be rented from manufacturers for on-site evaluation. Generally areal removal rates are lower in prototype RBC units than in bench scale units because the average mixing intensity, surface shear and exposed liquid film thicknesses are different. At the present time, suitable scale-up equations based on fundamental principles are not available to relate bench scale and prototype RBC performance. While such expressions may be developed in the future, only a carefully designed testing program that includes consideration of the elements described above can help prevent unpleasant "surprises" following construction.

CONCLUSIONS

The RBC process has many applications for the treatment of wastewaters. Rapidly escalating power costs make the process even more promising for the long term, cost effective treatment of industrial wastewaters. However, because of unique wastewater characteristics, gross traditional parameters such as BOD_5 and suspended solids alone are inadequate for assessing RBC treatment for many industrial wastewaters. Only complete chemical characterization studies and careful pilot plant evaluation programs can prevent subsequent process failures and allow owners to obtain the full benefits potentially available from the RBC process.

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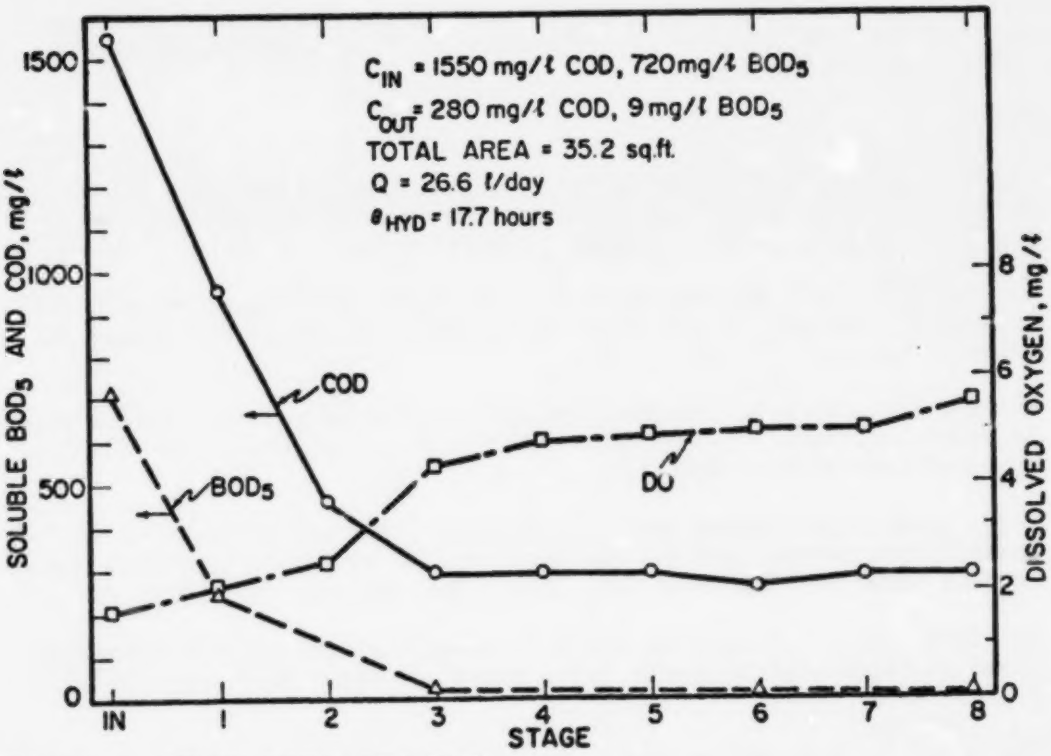

FIGURE I BENCH SCALE RBC DATA

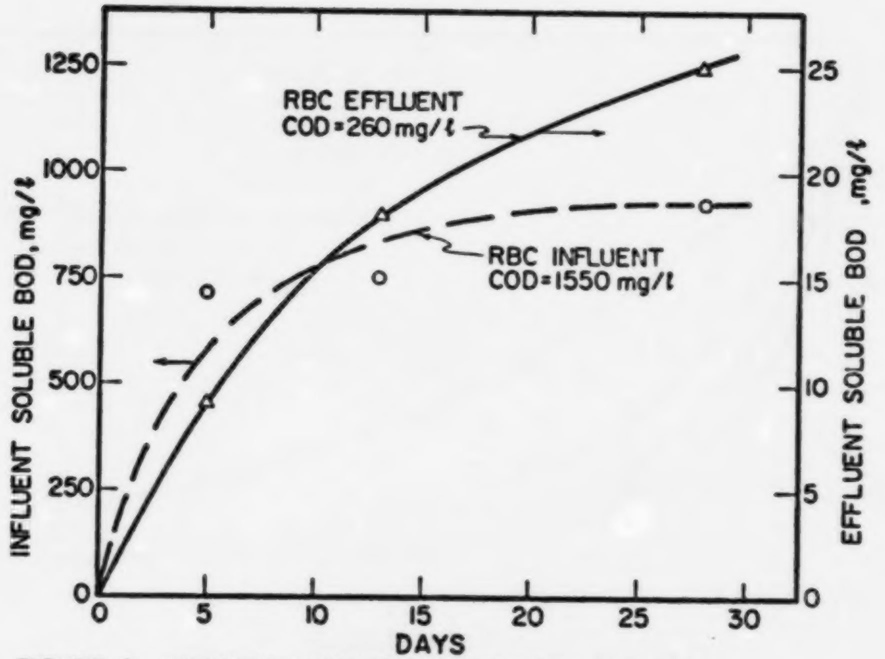

FIGURE 2 INFLUENT AND EFFLUENT BOD AND COD COMPARISONS

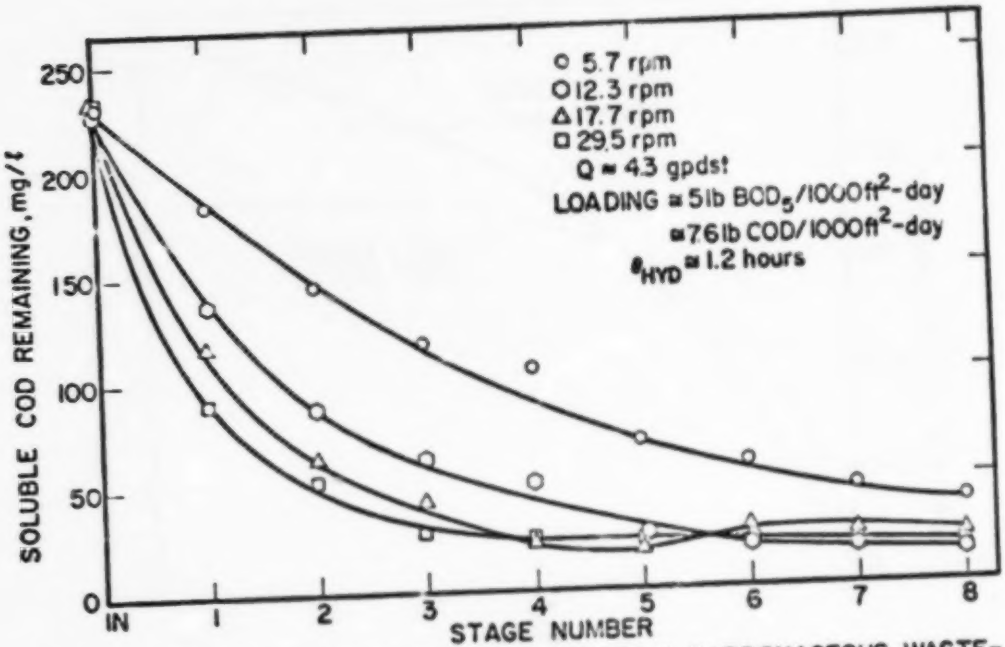

FIGURE 3 BENCH SCALE RBC TREATMENT FOR A CARBONACEOUS WASTEWATER

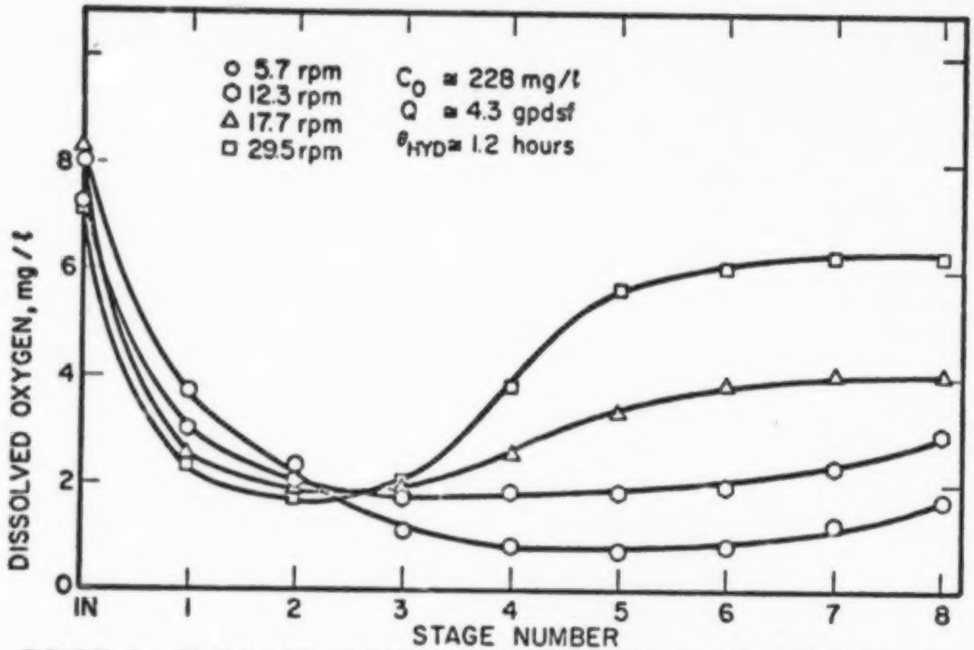

FIGURE 4 DISSOLVED OXYGEN PROFILES FOR A CARBONACEOUS WASTEWATER

ANALYSIS AND DESIGN OF ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

The basic mechanisms controlling the performance of rotating biological contactors in removing soluble substrate are twofold: (1) the transfer of substrate and electron acceptor (such as oxygen or nitrate) to the biomass attached to the rotating surfaces, and (2) the rate of biological growth and/or reaction occurring in the biomass. To properly understand the functioning of the RBC process, to confidently scale up pilot or bench scale data to plant design, and to get a first cut evaluation of process modifications on system performance, it is desirable to combine the mass transfer and kinetic mechanisms with the system hydraulics.

Our research efforts[1,2,3] over the last three years have incorporated the above first principles modeling approach to analyze and design RBC systems. This paper will summarize the results of this effort.

The approach taken is to summarize the model characteristics for both carbon oxidation and nitrogen removal with sufficient validation presented. The major thrusts of the paper are to quantify the importance of the various removal mechanisms acting in the system, indicate a design approach using the

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model for cabon oxidation, and evaluate the effect of wastewater characteristics and process modifications on system performance.

MODEL DEVELOPMENT

The RBC model is applicable to soluble organic carbon (BOD or COD) removal, nitrification, and soluble nitrogen removal. It consists of material balance equations that are numerically integrated to obtain concentrations of the carbonaceous and nitrogenous substrates in the liquid film and biofilm attached to the rotating media as well as concentrations in the mixed tank and effluent liquid from each stage. For the mass transfer process, the elements included in this model are oxygen and substrate transfer through liquid and biofilms, ammonia stripping, and subsurface aeration for an air drive system. All mass transfer coefficients in the model are related to system operating conditions. Reaction rates are also simultaneously related to electron donor (substrate) and electron acceptor (oxygen or nitrate) concentrations requiring no prior assumptions as to which component is controlling system performance.

For the model framework, the media is assumed to consist of flat plates divided into four stationary pie-shaped sectors as shown in Figure 1.

A stationary liquid film is utilized and gradients of all components through the biofilm are obtained by dividing the biofilm into completely mixed layers as shown in Figure 2. Since the reaction rate of the autotrophs is slower than the heterotrophs, deeper penetration of components exist in nitrification stages compared to those with carbon oxidation. Liquid film and biofilm are transported through stationary sectors at volumetric flow rates Q_L and Q_F respectively, both a function of rotational speed and layer thickness.

Mass Balance Equations

For each RBC stage, 3 types of mass balance equations are required: (1) tank equations, (2) biofilm equations and (3) liquid film equations. Referring to Figures 1 and 2, the non-steady state equation for the tank for each of the components considered in the model is:

Tank:

$$Q(S^T - S^{TO}) - K_S A \sum_{j=1}^{N_W} (S^T - S_j^1) + R_S V + K_L aV(S^* - S^T)$$

$$+ Q_L (S^{L4} - S^T) = V \frac{dS^T}{dt} \qquad (1)$$

All terms in the above equation are defined in the nomenclature section of the paper.

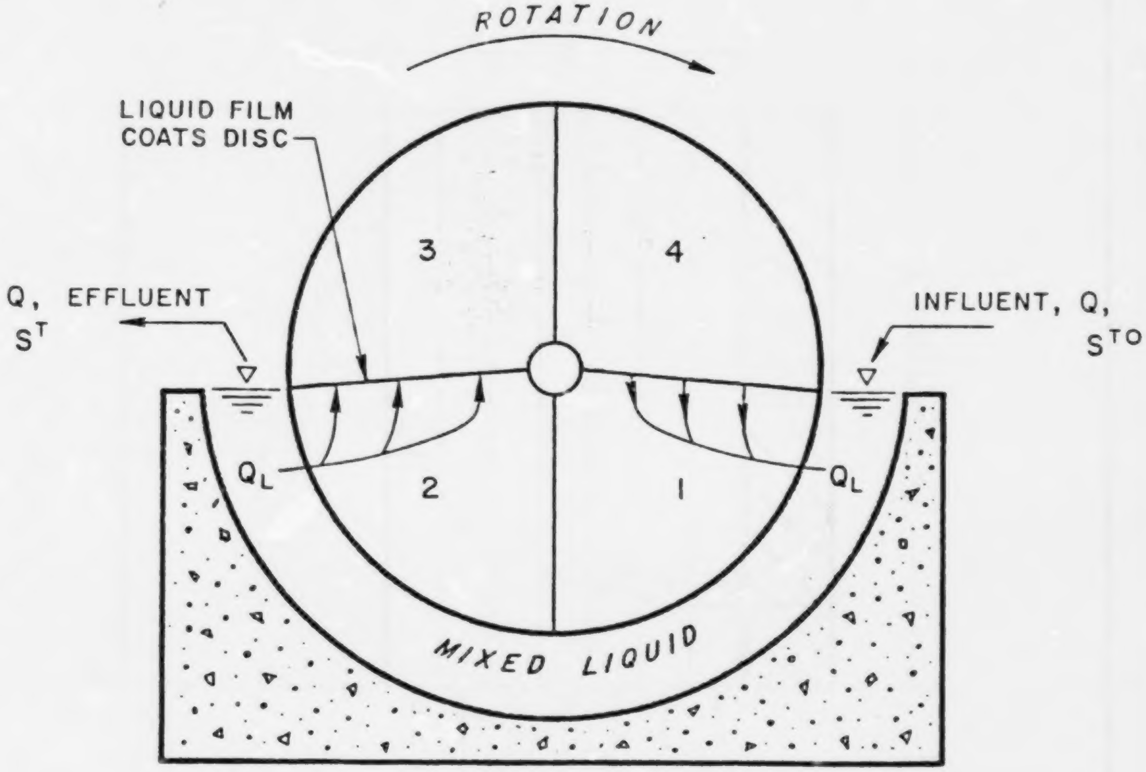
ROTATION
LIQUID FILM
COATS DISC
3
4
Q, EFFLUENT
S^T
INFLUENT, Q,
S^{TO}
Q_L
2
1
Q_L
MIXED LIQUID

FIGURE I
SKETCH OF SECTORS IN THE RBC MODEL

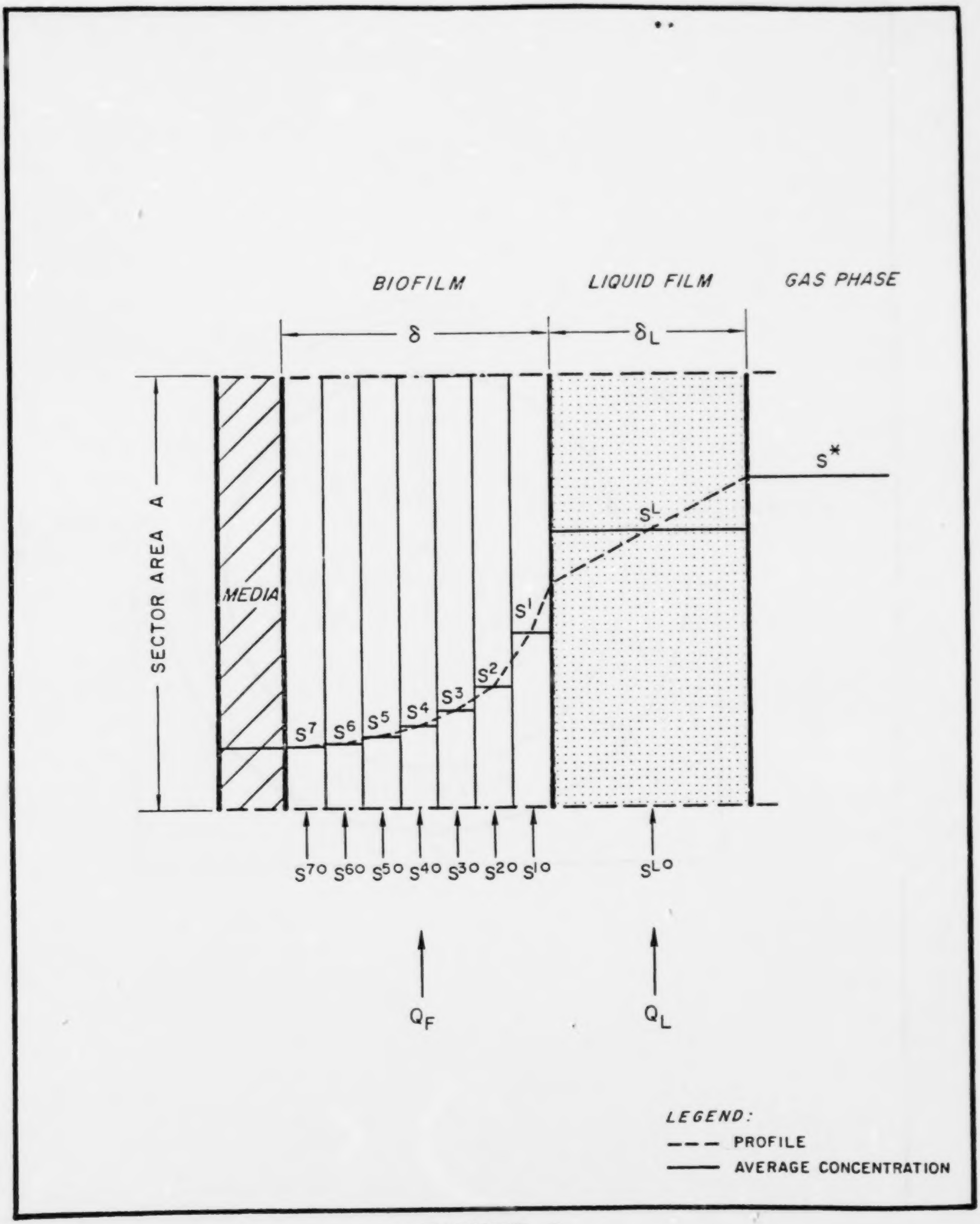
BIOFILM
LIQUID FILM
GAS PHASE
δ
δL
SECTOR AREA A
MEDIA
S*
SL
S1
S2
S3
S4
S5
S6
S7
S70
S60
S50
S40
S30
S20
S10
SLO
QF
QL
LEGEND:
PROFILE
AVERAGE CONCENTRATION

FIGURE 2
BIOFILM SCHEMATIC DIAGRAM

The mass balance equation for the tank not only includes the (1) advection terms and (2) mass transfer to the biofilm, which normally contribute the major portion of the overall concentration change in the tank, but also (3) reaction due to organisms (VSS) suspended in the tank mixed liquid as well as (4) gas transfer in the tank when an air drive model is used. To numerically integrate the above equation and solve for the tank concentration, S^T, requires knowledge of the concentration in the first biofilm layer, S^1, as well as that existing in the liquid film, S^{L4}, just before it reenters the tank. For simplicity, no separate liquid film is assumed to exist in the submerged sectors. Thus K_S is an overall transfer coefficient from the tank contents to the first biofilm layer in the submerged sectors.

To obtain the biofilm concentration, S^1, in the first layer for each sector, a material balance on an infinitesimal slice of biofilm yields the following differential equation:

Biofilm:

$$D_s \frac{\partial^2 S}{\partial y^2} + \frac{Q_F}{A\delta} (S^o - S) + R_S = \frac{\partial S}{\partial t} \tag{2}$$

In the above equation, the diffusion term relates the concentration in the layer being analyzed to its adjacent layers in the same sector. Transport of mass by diffusion to adjacent sectors is assumed negligible compared to the advective transport term. For the internal biofilm layers, the above equation is used for each component. At the liquid-biofilm interface and the media - (or dormant biofilm) biofilm interface, the following boundary conditions apply:

Liquid-Biofilm interface below water line:

$$-D_S \frac{\partial S}{\partial y} = K_S(S^T - S) \quad @ \; y = 0$$

Liquid-Biofilm interface above water line:

$$-D_S \frac{\partial S}{\partial y} = K_S(S^L - S) \quad @ \; y = 0$$

Media-Biofilm interface:

$$-D_S \frac{\partial S}{\partial y} = 0 \qquad @ \; y = \delta$$

The first two boundary conditions couple the biofilm concentration with the tank and liquid film concentrations. The last boundary condition results since the transport into the media or the dormant biofilm layer is zero. This effectively limits all reaction to the active biofilm thickness used in the model.

To finish the equation set, a material balance equation for the liquid film in each sector above the water line is required as follows:

Liquid Film:

$$\frac{Q_L}{A\delta_L}\ (S^{LO} - S^L) - \frac{K_S}{\delta_L}(S^L - S) + \frac{K_L}{\delta_L}(S^* - S^L) + R_S = \frac{\partial S^L}{\partial t} \qquad (3)$$

The above equation contains mass transfer to the biofilm which applies for all parameters and also mass transfer from the gas phase to the liquid film which applies only for the gaseous components, oxygen and non-ionized ammonia.

Reaction and Growth Kinetics

Biological reaction is based on the mixed liquor volatile suspended solids concentrations existing in the tank, liquid film, and biofilm. Two organism species are utilized: the heterotrophs using organic carbon as the substrate and the autotrophs using ammonia as substrate.

In the RBC model, five reacting components are simultaneously considered: BOD, O_2, NO_3-N, NH_4^+-N, and soluble organic nitrogen (SON). Figure 3 indicates schematically the transport and reaction mechanisms incorporated in the model. Interphase transport of both oxygen and ammonia between the gas phase and the liquid film occurs while the disc is rotating above the liquid surface. From the liquid film, the five diffusing species penetrate by molecular diffusion into the biofilm where biological reaction takes place.

Oxygen is used for organic carbon (BOD) oxidation, endogenous respiration, and nitrification.

Soluble organic nitrogen is hydrolyzed by both autotrophic and heterotrophic species. Initial model verification using only heterotrophic species for organic nitrogen hydrolysis proved inadequate. A review by Painter[4] indicates that hydrolysis by both species is possible although little data is available for substantiation. Oxidation of ammonia nitrogen to nitrate by the autotrophs provides energy for autotrophic growth. For model simplicity, intermediate nitrite production is ignored. Although endogenous respiration of autotrophs is provided for in the model framework, for simplicity and ease of mathematical computation, an endogenous rate coefficient of zero was utilized with a net cell yield value for autotrophic growth.

In the inner biofilm layers which have low oxygen concentrations, denitrification reactions occur both for BOD removal and endogenous respiration. Except for organic nitrogen hydrolysis which is independent of other reactions, the rates of reaction of all species are related stoichiometrically to the rates of heterotrophic and autrotrophic growth. Table 1 presents the substrate and electron acceptor rate equations for the five species of concern. Note that all rate equations are written as production equations for uniformity.

The growth rate equations used in Table 1 are defined in Table 2. The net growth rates of both species are delineated with respect to growth and endogenous respiration rates using oxygen or nitrate for ease of use in the substrate and electron acceptor equations. Michaelis kinetics are used for both electron donor and acceptor allowing either component to control. In

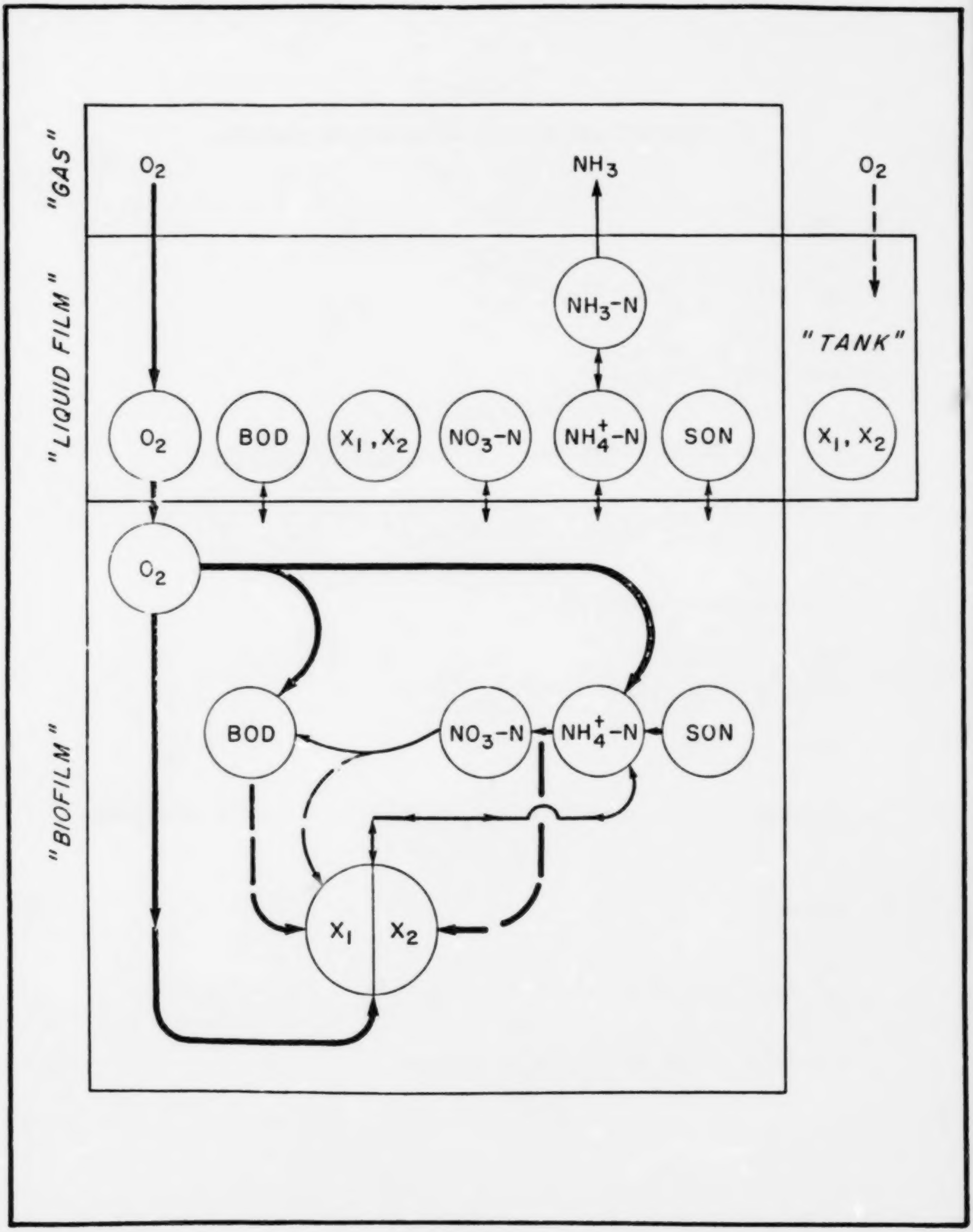
"GAS"
"LIQUID FILM"
"BIOFILM"
"TANK"
O_2
NH_3
O_2
NH_3-N
O_2
BOD
X_1, X_2
NO_3-N
NH_4^+-N
SON
X_1, X_2
O_2
BOD
NO_3-N
NH_4^+-N
SON
X_1
X_2

FIGURE 3

NITRIFICATION REACTION SCHEMATIC FOR AN RBC SYSTEM

TABLE 1

SUBSTRATE AND ELECTRON ACCEPTOR RATE EQUATIONS

1. BOD, S_1

$$\frac{dS_1}{dt} = R_1 = \frac{-1}{Y_1} [RX_{13} + RX_{14}]$$

where: $Y_1 = \frac{(f_1 - a_1)}{f_3}$;

$a_1 = \Delta O_2/\Delta$ Soluble Substrate $\quad f_1 = \frac{\Delta \text{ Soluble COD}}{\Delta \text{ Soluble Substrate}}$;

$f_3 = 1.42$ mg O_2/mg X

2. Total Ammonia Nitrogen, S_2

$$\frac{dS_2}{dt} = R_2 = \frac{-RX_2}{Y_2} - f_2 [RX_1 + RX_2] - R_5$$

where: Y_2 = net

Autotrophic cell yield = 0.17 mg X_2/mg NH_4^+-N ; $f_2 = 0.124$ mg NH_4^+-N/mg X

3. Oxygen, S_3

$$\frac{dS_3}{dt} = R_3 = -\frac{a_1}{Y_1} RX_{13} - \frac{a_2}{Y_2} RX_2 - f_3 [EX_{13} - RX_2]$$

where: $a_2 = 4.57$ mg O_2/mg NH_4^+-N oxidized

TABLE 1 (Continued)

4. Nitrate N, S_4

$$\frac{dS_4}{dt} = R_4 = \frac{RX_2}{Y_2} - f_4 \left[\frac{a_1}{Y_1} RX_{14} + f_3 EX_{14}\right]$$

where: $f_4 = 0.35$ mg NO_3-N/mg O_2

5. SON, S_5

$$\frac{dS_5}{dt} = R_5 = -k_5 X \frac{S_5}{S_5+S_{5m}} \frac{S_3}{S_3+S_{3m}}$$

TABLE 2

GROWTH RATE EQUATIONS

1. Net Heterotrophic Growth Rate, RX_1

$$RX_1 = RX_{13} + RX_{14} - EX_{13} - EX_{14}$$

a) $RX_{13} = \mu_1 X_1 \frac{S_1}{S_1+S_{1m}} \frac{S_3}{S_3+S_{3m}}$, Growth Rate Using O_2

b) $RX_{14} = \mu_1 X_1 \frac{S_1}{S_1+S_{1m}} \frac{S_{3m}}{S_3+S_{3m}} \frac{S_4}{S_4+S_{4m}}$, Growth Rate Using NO_3-N

c) $EX_{13} = b_1 X_1 \frac{S_3}{S_3+S_{3m}}$, Endog. Resp. Rate Using O_2

d) $EX_{14} = b_1 X_1 \frac{S_{3m}}{S_3+S_{3m}} \frac{S_4}{S_4+S_{4m}}$, Endog. Resp. Rate Using NO_3-N

2. Net Autotrophic Growth Rate, RX_2

$RX_2 = \mu_2 X_2 \frac{S_2}{S_2+S_{2m}} \frac{S_3}{S_3+S_{32m}}$, Growth Rate Using O_2

the denitrification reactions, an additional oxygen Michaelis type term (S_{3m}/S_3+S_{3m}) is included which has the effect of allowing denitrification to occur when the oxygen concentration is reduced to values near or below the oxygen Michaelis constant. Since nitrification is more significantly affected by low oxygen concentrations, different oxygen Michaelis values are used for substrate oxidation and nitrification.

To solve the above equations, the organism distribution between heterotrophs and autotrophs must be determined when nitrification is present. This is accomplished by setting the ratio of the organism concentrations in each stage equal to the ratio of their growth rates in the biofilm for that stage. Thus:

$$\frac{X_2}{X_1} = \frac{\sum_{1}^{n} RX_2}{\sum_{1}^{n} RX_1} \qquad (4)$$

where n = number of biofilm segments in all sectors of a stage.

In addition to the above reaction terms, the ionized ammonia fraction must be estimated for use in the ammmonia stripping equation as follows:

$$S_{2N} = f_N\, S_2 \qquad (5)$$

where $f_N = 1/1 + 10^{pK_a-pH}$ = non-ionized fraction and $pK_a = 10.05 - 0.032\, t(^\circ C)$ (Ref. 5).

Mass Transfer Coefficients

The mass transfer coefficients for oxygen and substrate as well as liquid film thickness were obtained using the theory of withdrawal of a flat plate.

Levich[6] analyzed the situation of the film substrate entrained on a flat plate vertically withdrawn from a quiescent liquid and deduced that the film thickness h_o could be represented by the following equation:

$$h_o = 0.944\,\frac{(\mu v)^{2/3}}{\sigma^{1/6}(\rho g)^{1/2}} \qquad (6)$$

Assuming that the physical properties of the wastewater will not differ appreciably from pure water at 20°C, equation (12) reduces to:

$$h_o = 6.85\, v^{2/3} \qquad (7)$$

where v is the withdrawal velocity.

Due to irregularities in the biofilm surface, the retained liquid film will be thicker than that obtained on a flat plate. Williamson and McCarty[7] measured surface irregularities and found depths ranging between 16 and 30 microns. Based on these measurements it was decided to add 25 microns to h_o to obtain the liquid film thickness δ_L. Since withdrawal velocity in a rotating system is a function of radial position, an average withdrawal velocity at the centroid of mass (2/3 of media radius) was employed.

The mass transfer coefficients are related to δ_L through the equations:

$$K_s = \frac{D'_s}{\delta_L/2} \quad , \quad K_L = \frac{D'_o}{\delta_L/2} \tag{8}$$

where, D'_o and D'_s are diffusivities of oxygen and substrate in water, and $\delta_L/2$ is the diffusion path length from the average concentration to the interface concentration.

In developing the oxygen transfer equation from air to the liquid film, the saturation concentration for oxygen, S_3, is assumed to remain constant requiring the partial pressure in the gas phase to remain constant. If one assumes that the air enters the internal media passages as it leaves the tank liquid and remains there until discharged as the media reenters the tank, the decrease in O_2 partial pressure is negligible for any reasonable O_2 transfer rate. Thus the assumption of a constant S_3 is valid. This is due to the relatively low solubility of oxygen and high partial pressure (20.9%) of oxygen in air.

However, when evaluating ammonia stripping the reverse is true, ammonia being a highly soluble gas with a partial pressure of zero typically existing in the atmosphere. If one utilizes the previous stagnant film techniques to calculate K_{L2} with the saturation value $S_2 = 0$, then the amount of ammonia stripping occurring is greatly exaggerated, since the gas phase composition would require continuous instantaneous replacement in the internal media voids.

Since gas flow rate appears to be the major factor affecting ammonia stripping, the following approach using a saturated gas flow was utilized. A gas phase flow rate is calculated as a function of rotational speed similar to the liquid film flow rate. The ammonia partial pressure in the gas phase is assumed to be in equilibrium with the non-ionized ammonia concentration in the liquid film of the last aerated sector prior to reentering the tank contents. In equation 3, for the liquid film, the amount of ammmonia transferred from the gas phase becomes:

$$\frac{K_{L2}}{\delta_L}(S_2^* - f_N S_2) = -\frac{K_{L2} f_N S_2}{\delta_L} = -\frac{M_2 p_2 G}{RT\, A_A \delta_L} \tag{9}$$

Since $p_2 = f_N S_2/H_2$ from Henry's Law and the gas flow can be given by $G = \overline{V}_g \omega = R_H A_T \omega$, then the ammonia transfer coefficient above the water line is:

$$K_{L2} = \frac{M_2 \; R_H A_T \omega}{RT \; H_2 A_A} \qquad (10)$$

Subsurface Gas Transfer - Air Drive

The additional aeration occurring for the air drive system can be evaluated using similarities to diffused aeration systems. For diffused aeration, the oxygen transfer coefficient is related to gas flow and transfer efficiency by the following:

$$K_L a_{20} V = \frac{E_o}{100} \frac{1.04 \; G_o}{C_{s20} \; 8.34} \; [=] \; \frac{MG}{hr} \qquad (11)$$

where G_o = gas flow, scfm, E_o = % absorption efficiency. The above equation is related to standard conditions of 20°C. Temperature scaleup is typically given by:

$$K_L a_t = K_L a_{20} \; 1.02^{t(oC)-20} \qquad (12)$$

For ammonia stripping, the assumption of saturated gas flow is again made which provides the maximum quantity of ammonia removal possible.

Numerical Evaluation - Method of Solution

Although the tank, biofilm, and liquid flow mass balance equations (1-3) are presented as time variable or non-steady state equations, they were solved as steady-state equations for the purposes of the model calibrations presented herein. Under the assumption of steady-state, the mass-balance equations reduce to a set of simultaneous non-linear algebraic equations, which may be written as:

$$\begin{aligned}
a_{11}s_1 + a_{12}s_2 + a_{13}s_3 + \dots + a_{1n}s_n &= b_1 \\
a_{21}s_1 + a_{22}s_2 + a_{23}s_3 + \dots + a_{2n}s_n &= b_2 \qquad (13) \\
&\vdots \\
a_{n1}s_1 + a_{n2}s_2 + a_{n3}s_3 + \dots + a_{nn}s_n &= b_n
\end{aligned}$$

or in standard matrix notation:

$$[A](s) = (b)$$

where:

s_i is the (unknown and to be solved for) concentration in compartment i

a_{ii} is comprised of the advective and mass transfer terms, and the reactive and kinetic terms involving s_i

a_{ij} is comprised of the advective and mass transfer terms, and the reactive and kinetic terms between compartments i and j, involving s_j

b_i is the effective driving force due to appropriate influent conditions and boundary conditions,

where a compartment is defined to be a particular component (BOD, NH_3, O_2, etc.) in a particular medium (tank, biofilm, liquid film), in a particular stage of the rotating biological contactor.

Due to the non-linear nature of the mass balance equations and the number of equations involved (61 cells x 5 components = 305 equations and unknowns/stage), use of conventional matrix inversion to solve for the unknown is ruled out. Instead, a modified form of Gauss-Seidel iteration is employed. The Gauss-Seidel iteration method involves rewriting each of the equations in (13) in terms of the unknown as follows:

$$\begin{aligned} s_1 &= \frac{b_1 - a_{12}s_2 - a_{13}s_3 - \dots a_{1n}s_n}{a_{11}} \\ s_2 &= \frac{b_2 - a_{21}s_1 - a_{23}s_3 - \dots a_{2n}s_n}{a_{22}} \\ &\vdots \\ s_n &= \frac{b_n - a_{n1}s_1 - a_{n2}s_2 - \dots a_{n\,n-1}s_{n-1}}{a_{nn}} \end{aligned} \tag{14}$$

Starting with an initial solution estimate for (s), the above equations are iteratively solved until the residual differences between two successive iterations is within a set tolerance. This promise may be shown as follows:

$$\begin{aligned} s_1^{k+1} &= \frac{b_1 - a_{12}s_2^k - a_{13}s_3^k - \dots a_{1n}s_n^k}{a_{11}} \\ s_2^{k+1} &= \frac{b_2 - a_{21}s_1^{k+1} - a_{23}s_3^k - \dots a_{2n}s_n^k}{a_{22}} \\ &\vdots \\ s_1^{k+1} &= \frac{b_n - a_{n1}s_1^{k+1} - a_{n2}s_2^{k+1} - \dots a_{n\,n-1}\,s_{n-1}^{k+1}}{a_{nn}} \end{aligned} \tag{15}$$

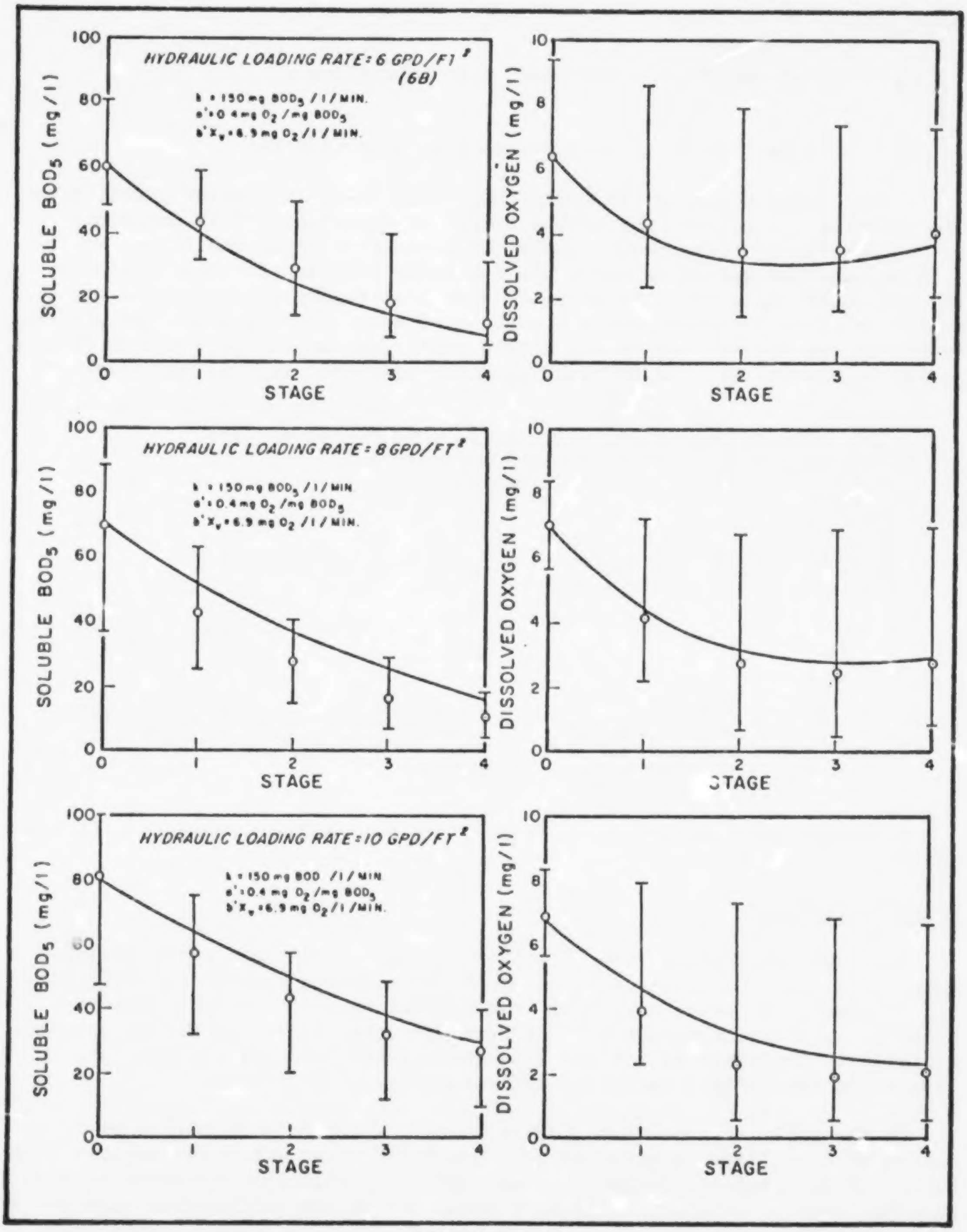

FIGURE 4

CARBONACEOUS MODEL CALIBRATION – ZERO ORDER KINETICS
PAPERMILL WASTEWATER

where:

s_n^k = concentration of compartment n at interation k

s_n^{k+1} = concentration of compartment n at iteration k+1

It was found necessary to modify the Gauss-Seidel iteration method after experiencing non-convergence of the system of equations. This non-convergence occurred in compartments where the concentration of one of the variables of interest was at or below the Michaelis constant for that variable. The non-convergence is to be expected when that occurs since this is the region in which the equations are most non-linear. This problem was overcome by taking the derivatives of the growth rate equation of Table 2 with respect to the dependent variable and modifying the equations in (15) as follows:

$$s_1^{k+1} = \frac{b_1 - a_{12}s_2^k - a_{13}s_3^k - \dots a_{1n}s_n^k + (d(ds_1^k/dt)/ds_1)s_1^k}{a_{11} + d(ds_1^k/dt)/ds_1}$$

$$s_2^{k+1} = \frac{b_2 - a_{21}s_1^{k+1} - a_{23}s_3^k - \dots a_{2n}s_n^k + (d(ds_2^k/dt)/ds_1)s_2^k}{a_{22} + d(ds_2^k/dt)/ds_2}$$

$$\vdots$$

$$s_n^{k+1} = \frac{b_n - a_{n1}s_1^{k+1} - a_{n2}s_2^{k+1} - \dots a_{n\,n-1}\,s_{n-1}^{k+1} + (d(ds_n^k/dt)/ds_n)s_n^k}{a_{nn} + d(ds_n^k/dt)/ds_n}$$

This modified Gauss-Seidel procedure was successfully convergent in all cases.

MODEL CALIBRATION

Numerous calibration runs[1,2] were conducted to evaluate model adequacy using pilot and full scale data for both carbonaceous removal and nitrification systems. To date, only data from mechanical drive systems have been used for calibration purposes. The approach taken is to set Michaelis coefficients in the range of typical literature values and utilize the data to obtain the remaining kinetic and stoichiometric coefficients.

In the early carbonaceous studies, both low and high Michaelis coefficients were utilized to simulate zero and first order substrate removal kinetics. Stage dissolved oxygen data was used to obtain the substrate removal coefficient, $k = \mu_1 X_1/Y_1$. Figure 4 shows one of the calibration runs for a papermill wastewater using zero order kinetics. First order kinetics provided a similar fit but with a wider range of kinetic coefficients when different wastewaters were analyzed.

The RBC model developed in this study provides concentrations within the biofilm which can be used, in conjunction with substrate and oxygen Michaelis constants, to determine active biofilm depth. This is illustrated in the biofilm profiles shown in Figures 5 and 6. Concentrations employed to construct these profiles were taken from the computer output of the 10 gpd/ft^2 (407 l/d/sq. m) verification run of Figure 4.

The kinetic equations used in the RBC model cause a reduction in BOD removal rate when either of the ratios, $S/(S+S_m)$ for BOD or oxygen drops below unity. For example, if $S_3 << S_{3m}$ in regions of the biofilm where $S_1/(S_1+S_{1m})$ is approximately equal to one, reaction is limited by a deficiency of oxygen. Examination of Figure 5 reveals that this is clearly the case in stage 1. Substrate concentrations are in excess of 4.3 mg BOD/1 throughout the biofilm. Since S_{1m} = 0.1 mg BOD/1, substrate will reduce removal rate at most by 2.3 percent. On the other hand, oxygen concentrations dropped well below S_{3m} = .001 mg/1 at biofilm depths in excess of 120 microns. For practical purposes, the active biofilm depth of stage 1 is 120 microns, and substrate removal in the stage is limited by a deficiency of oxygen. The reverse situation exists in stage 4. As can be seen in Figure 6, except for one or two points in the biofilm, the oxygen concentration is much greater than its Michaelis constant and has a negligible effect in reducing BOD removal rate. In stage 4, a deficiency of substrate within the biofilm is the rate limiting factor. The average value of $S_1/(S_1+S_{1m})$ at the 120 micron depth was 0.06, causing a 94 percent reduction in removal rate.

For the nitrification system, five coefficients must be evaluated from the data: μ_1, μ_2, k_5, a_1, and b_1. Soluble BOD_5 and oxygen data were used to estimate μ_1, a_1 and b_1, while μ_2 was obtained from ammonia data and k_5 from SON data. NO_3 data then served as a check on the system.

Figures 7 and 8 show calibration results for two data sets (on the Gladstone, Michigan municipal wastewater). In the runs, no attempt was made to apply a consistent temperature correction factor to the various coefficients. Instead the best coefficients which fit the data were chosen and then used to evaluate the temperature effect. Also in these runs, a yield coefficient of 0.38 was used for the autotrophs based on thermodynamic considerations. In the remaining runs a value of 0.17 was used for the net cell yield coefficient. In all figures the units for μ_1, μ_2, k_5 and b_1 are day^{-1} while the Michaelis constants are in units of mg/l.

A wide range of nitrogen concentrations occur in Figure 9 over the three days analyzed possibly due to digester supernatant being returned to the primary clarifier during the major portion of this run. It was not returned for the remaining runs. In some runs little BOD_5 removal was at times measured in the first stage, requiring higher influent BOD_5 values for the best fit of all stage data.

The variation of the four kinetic parameters obtained from the above calibrations as a function of temperature is shown in Figure 9. For all parameters, a significant temperature effect is seen. Using the following equation to describe temperature efffects,

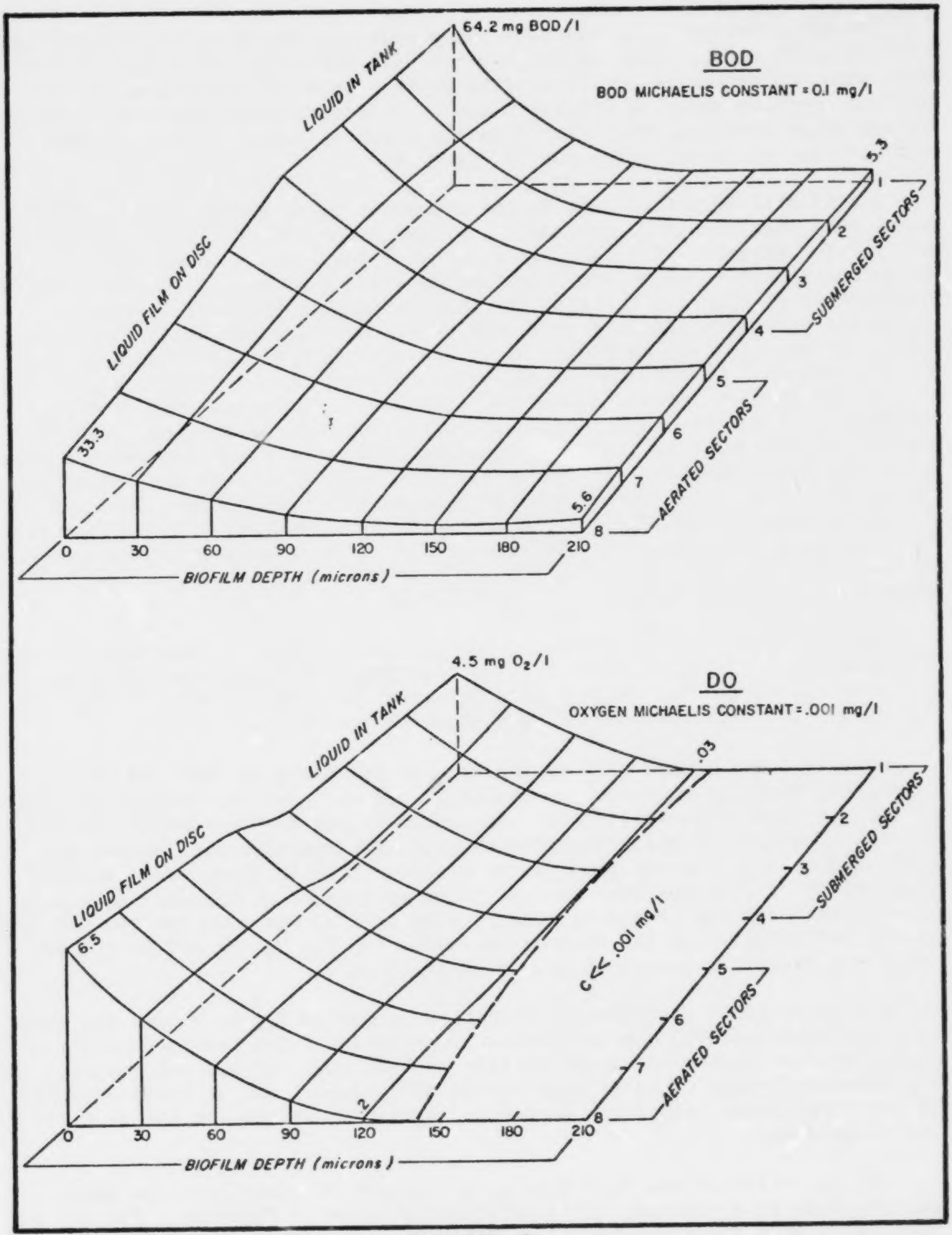

FIGURE 5
BIOFILM BOD AND DO PROFILES IN STAGE I OF PAPERMILL A PILOT PLANT AT 10 GPD/FT^2

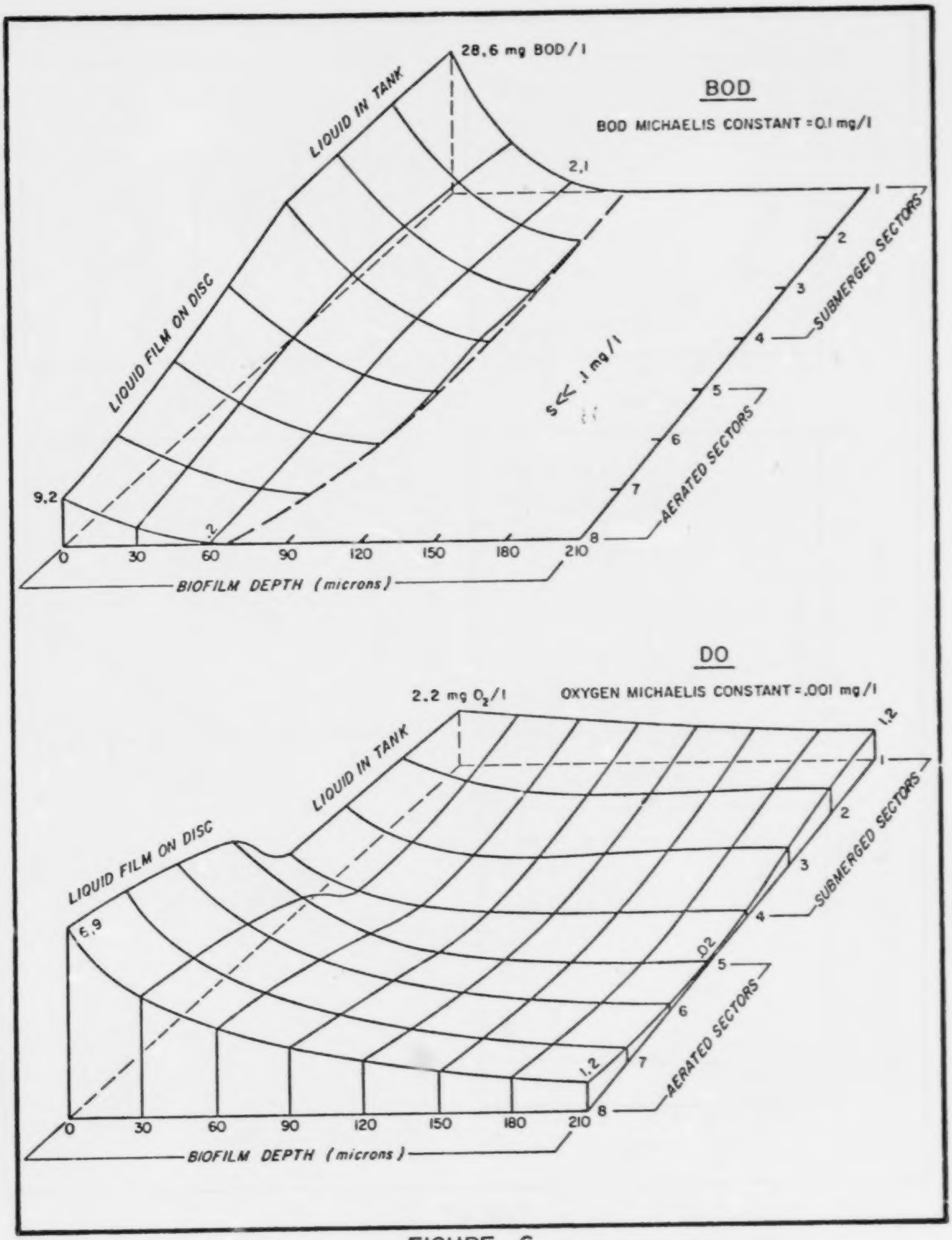

FIGURE 6

BIOFILM BOD AND DO PROFILES IN STAGE 4 OF PAPERMILL A PILOT PLANT AT 10 GPD/FT^2

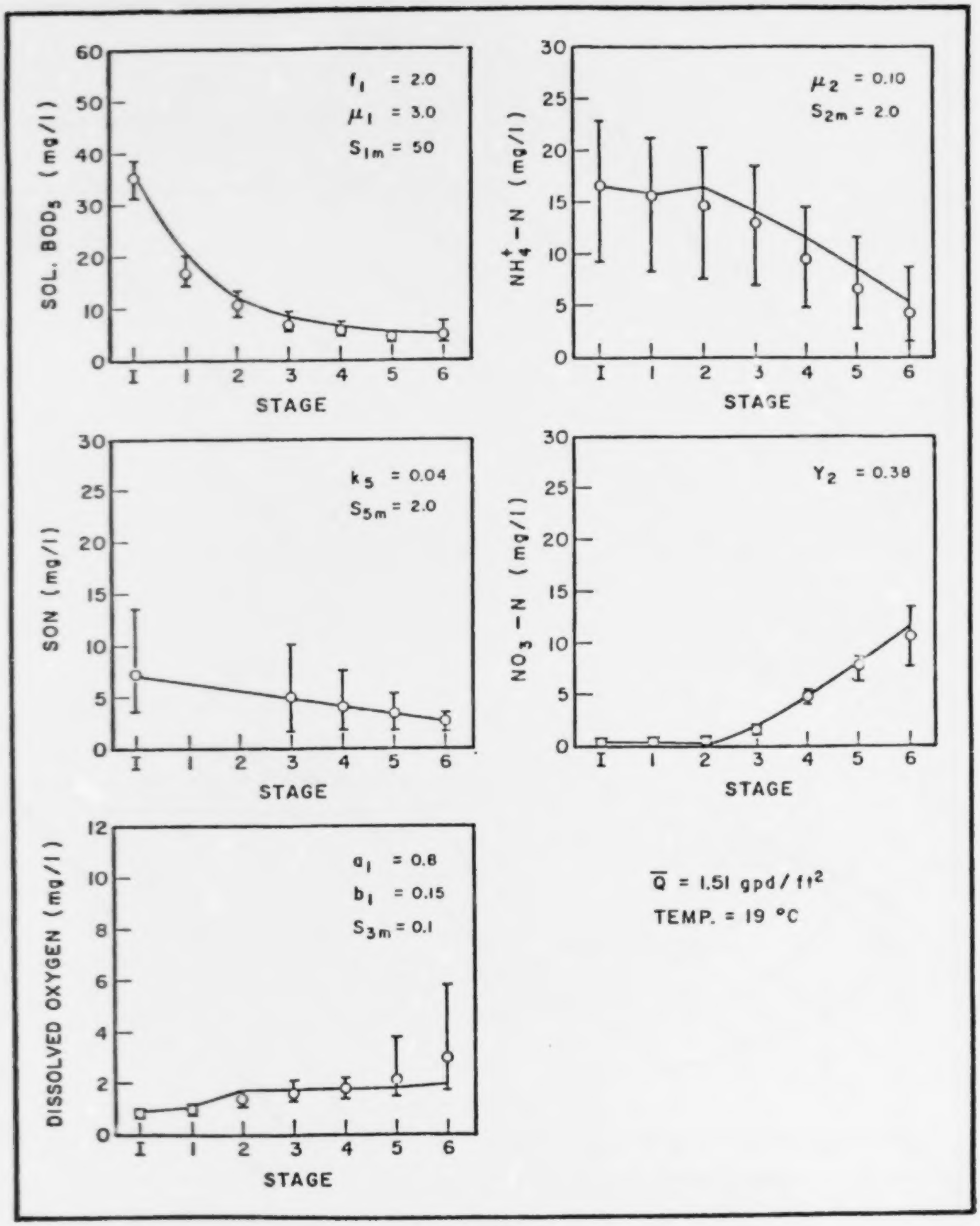

FIGURE 7

NITRIFICATION MODEL CALIBRATION – DATA SET I
MUNICIPAL WASTEWATER TEMPERATURE = 19°C

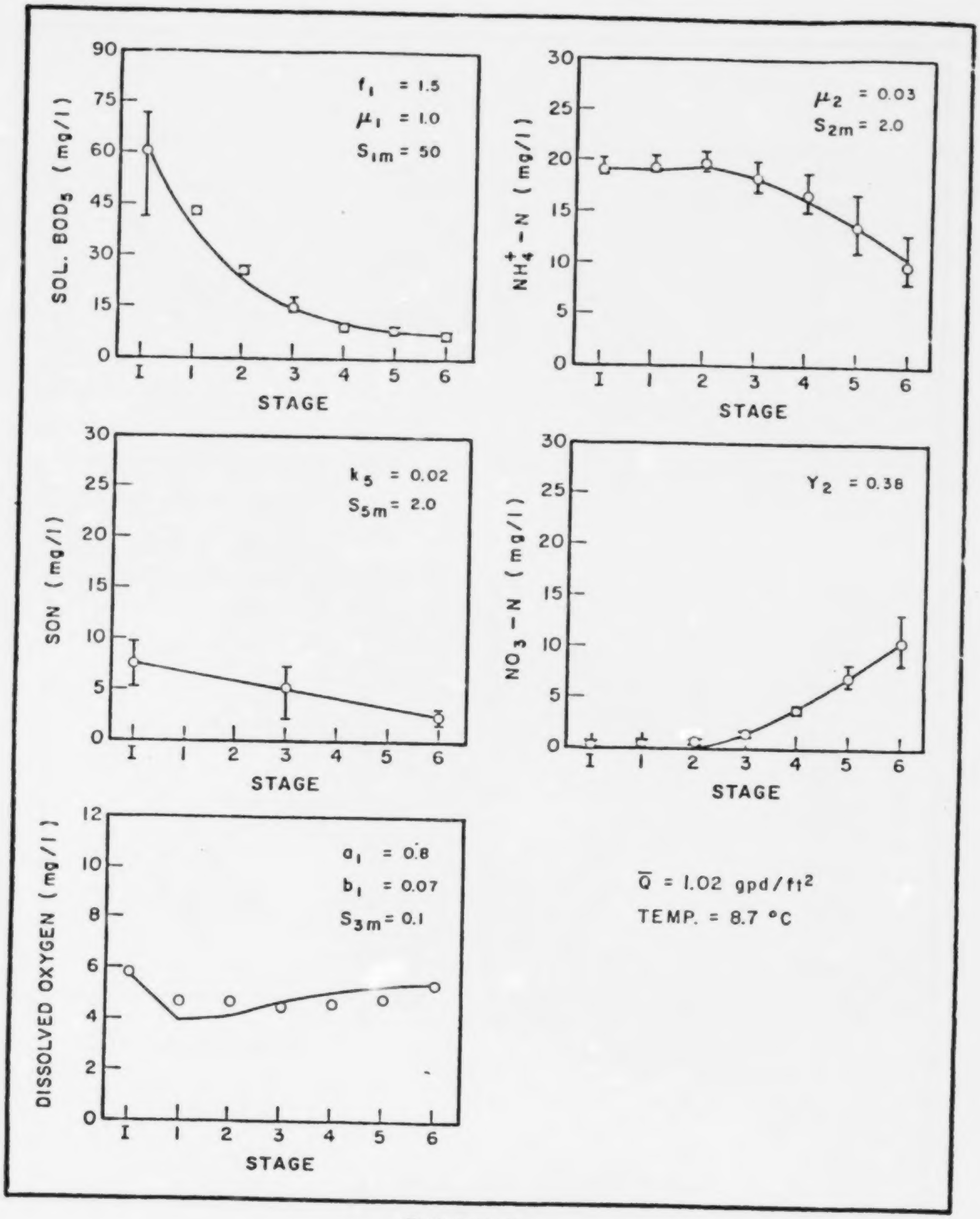

FIGURE 8

NITRIFICATION MODEL CALIBRATION – DATA SET 3
MUNICIPAL WASTEWATER TEMPERATURE = 8.7°C

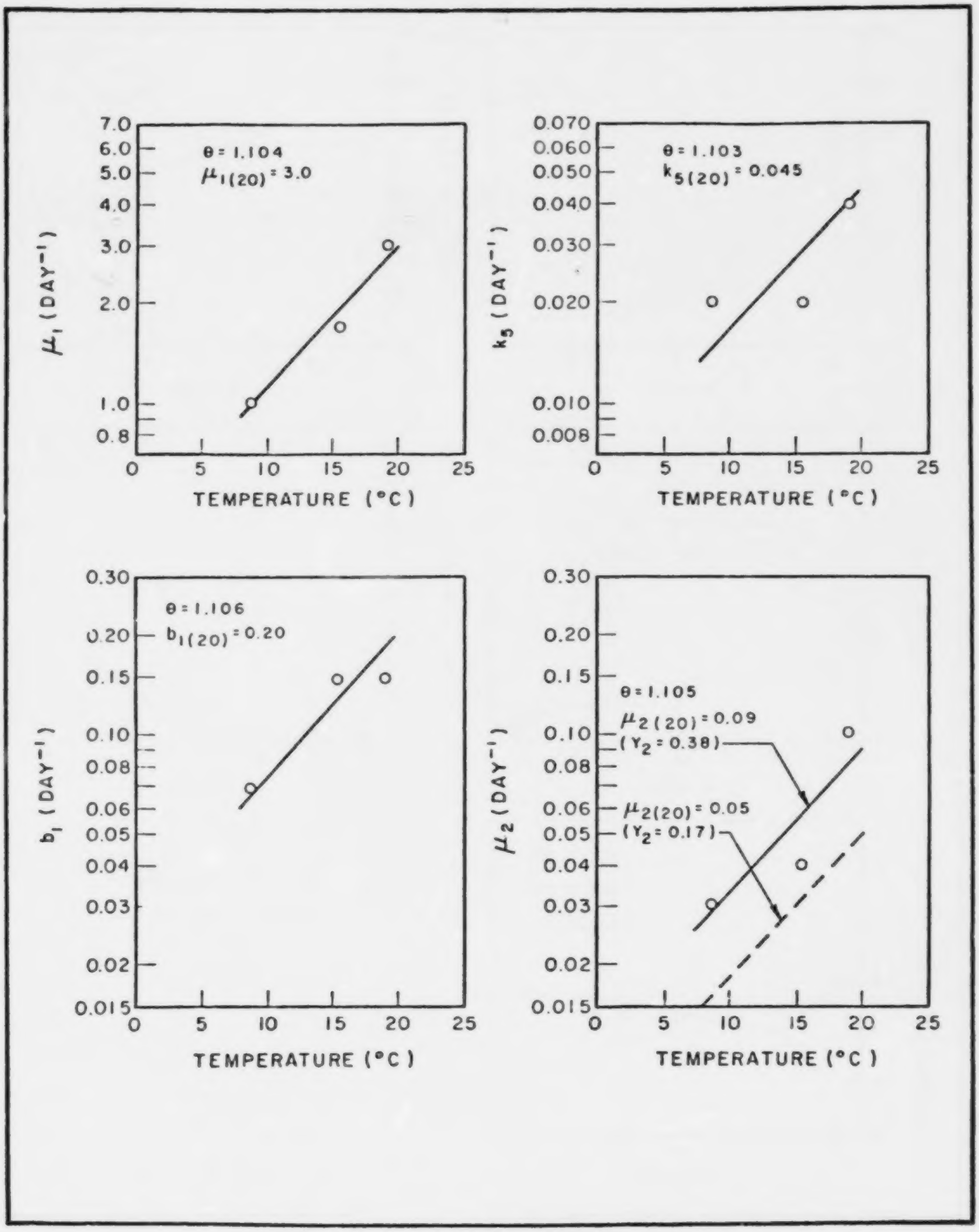

FIGURE 9

EFFECT OF TEMPERATURE ON KINETIC COEFFICIENTS

$$\mu_t = \mu_{20}\ \theta^{t\ (^{o}C)-20}$$

a θ value of 1.1 was chosen to represent all parameters. A value of this magnitude is typical for the nitrogen data but high for the BOD data. This may be due in part to the digester supernatant influent to the first data set. To obtain a greater degree of reliability for these temperature coefficients additional data sets should be analyzed. For the autotrophic growth rate, μ_2, a lower growth rate was required when the yield coefficient was reduced from 0.38 to 0.17 for the first three data sets; this lower value used for final verification runs[2].

Figures 10 and 11 illustrate the profiles resulting through the biofilm for the first and third stages for data set 1 (Figure 7). Sectors 2 and 4 only, the last sectors in the submerged and aerated portions of the disc, are presented to show the maximum differences in concentrations. For stage 1, sharp gradients in both BOD_5 and D.O. result through the biofilm. This is a result of the relatively high BOD reaction rate compared to the mass transfer rate. Much smaller gradients exist for the soluble organic nitrogen due to the lower reaction rate. The latter tends to be reaction limited not mass transfer limited as are BOD_5 and oxygen. The depth of oxygen penetration is significantly greater above the water line due to the higher D.O. values existing in the aerated sectors compared to the tank. Only 10% of the total biofilm depth of this stage (3180$_{\mu}$) is active in the aerated sectors and 3% in the submerged sectors indicating the large amount of excess biofilm carried in this stage. This is contrary to the value of 70% assumed by Clark, et al[8]. No nitrification occurs in this stage due to the heterotrophs completely outgrowing the autotrophs as seen later in Figure 12. In stage 3, nitrification is beginning since the BOD_5 has been reduced to a level allowing significant autotrophic growth to occur. The nitrate profile is very interesting in this stage since it peaks at an intermediate depth in the biofilm of 100-150 microns. Since nitrate is produced within the biofilm from nitrification, concentrations in the liquid film and tank are lower than the peak due to the concentration gradient required for diffusion out of the biofilm. In the internal portions of the biofilm where the oxygen concentration is near zero, significant denitrification occurs due to additional BOD_5 removal and heterotroph endogenous respiration. The ammonia gradient in the biofilm is slight, similar to the organic nitrogen, indicating the reaction controlled nature of nitrification except in the lower depths of biofilm where oxygen transport limits reaction. Oxygen penetration in the submerged sectors is about double that in the first stage due to the lower rates in the biofilm.

In Figure 12, the heterotrophs are seen to predominate entirely in the first two stages followed by a rapid decline with a proportional increase of the autotrophs in the latter stages. This is due entirely to the growth rates of the two species and the concentrations of BOD_5 and ammonia existing in the biofilm in these stages. The major mechanisms for TKN removal are also shown in Figure 12. In the first two stages, net growth is the only significant removal mechanism for TKN. In these first two stages growth removes approximately 2.3 mg/l TKN but the net removal is about one half that amount due to TKN production by endogenous respiration of the heterotrophs. The major fraction (~90%) of TKN removal is due to nitrification in the last four stages.

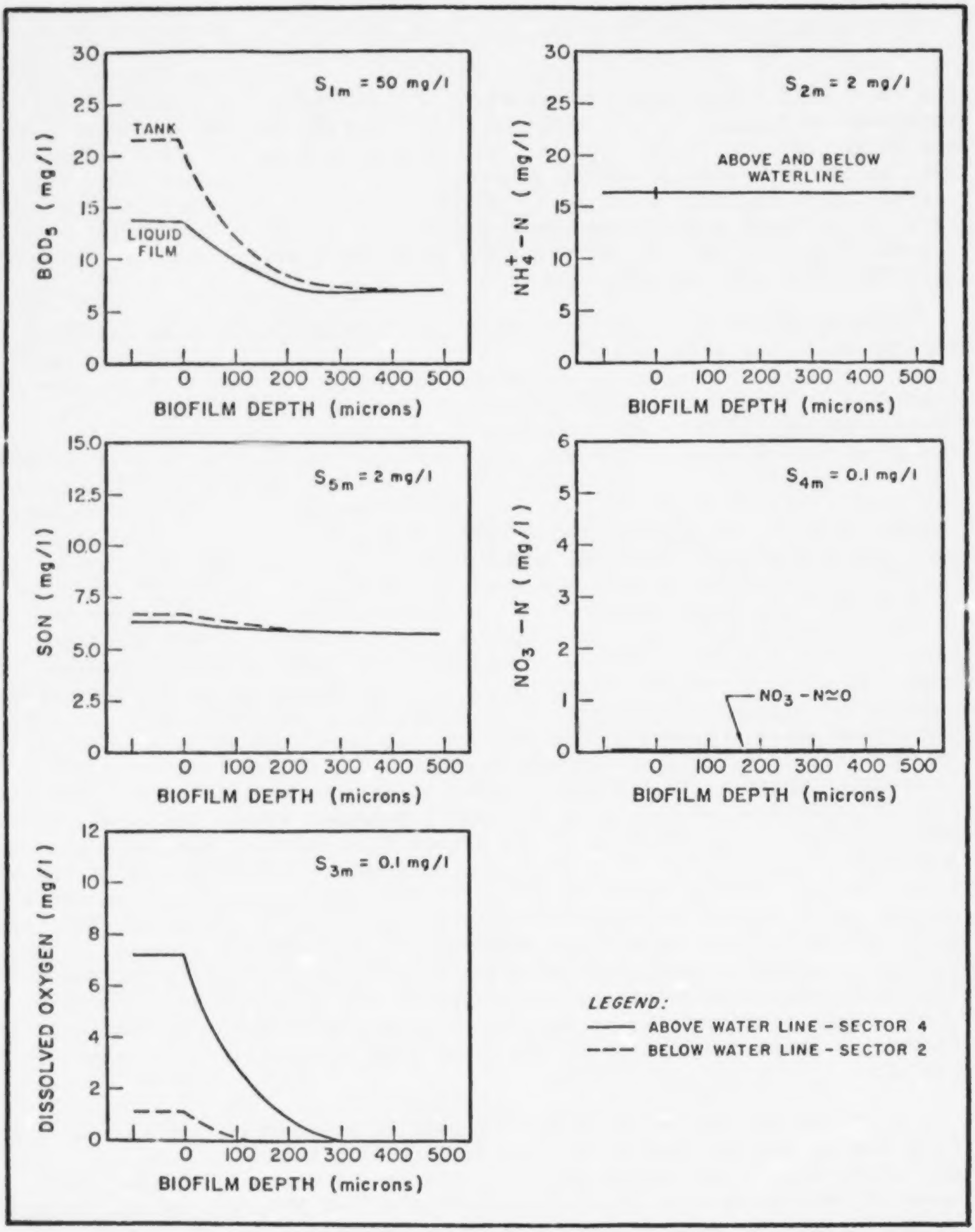

FIGURE 10
STAGE I CALCULATED PROFILES OF
BIOFILM CONCENTRATIONS – DATA SET I

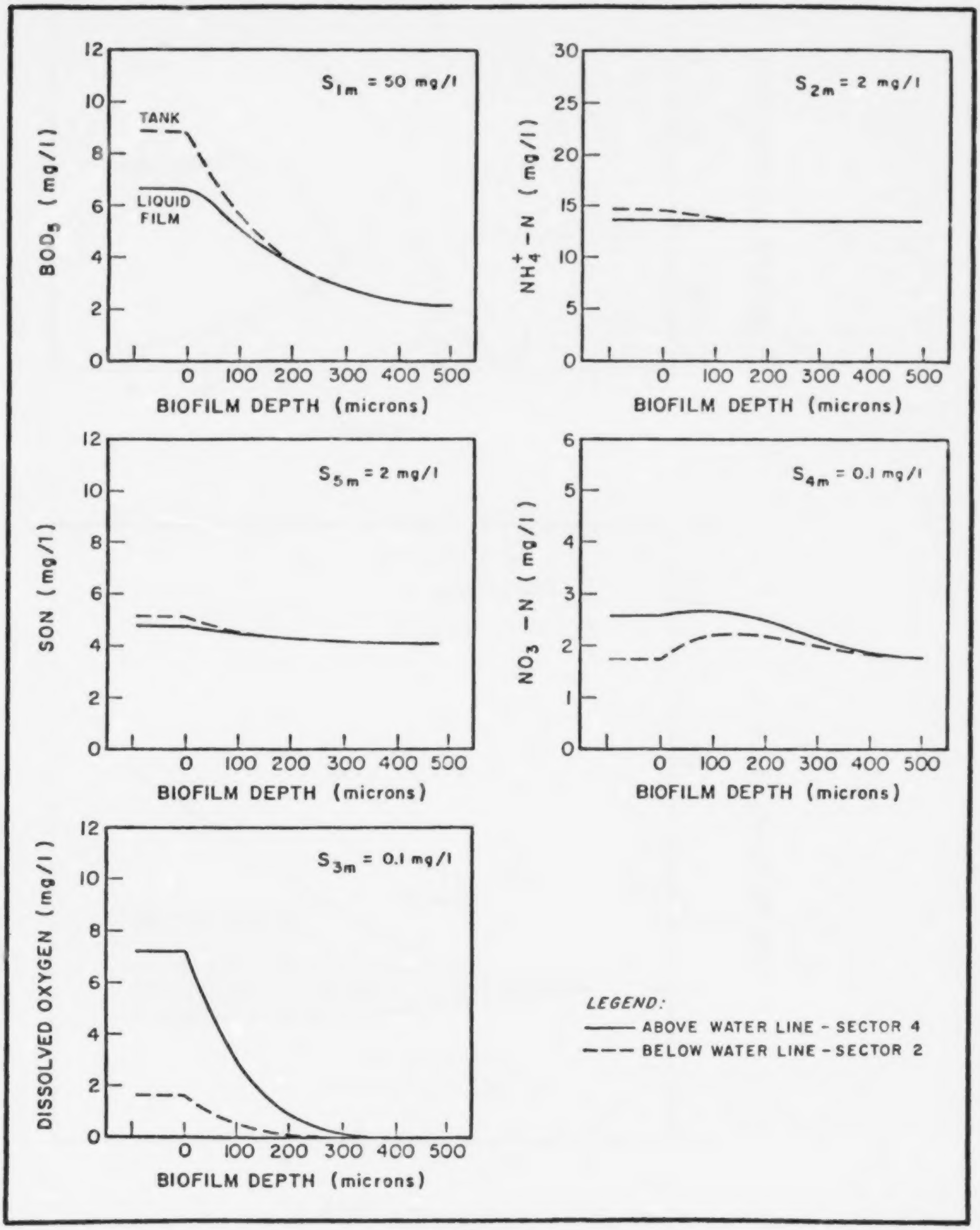

FIGURE 11
STAGE 3 CALCULATED PROFILES OF
BIOFILM CONCENTRATIONS - DATA SET 1

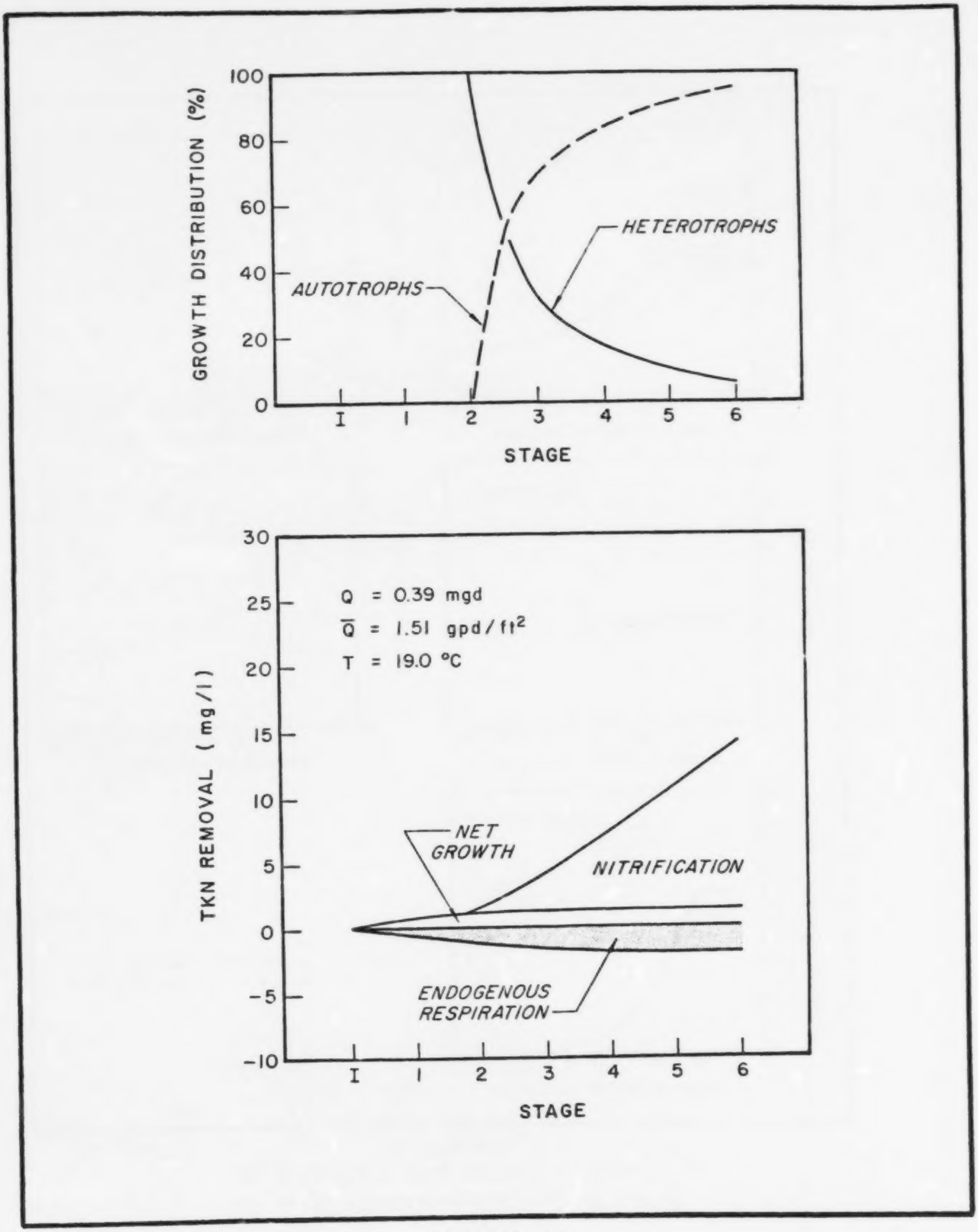

FIGURE 12

DISTRIBUTIONS OF GROWTH AND TKN REMOVAL FOR DATA SET I

The distribution of total soluble nitrogen removals for this run is shown in Figure 13. The major quantity of nitrogen lost from the system is due to denitrification which occurs mainly in the third and fourth stages when significant populations of both heterotroph and autotrophs exist. Some denitrification occurs in the first two stages due to 0.4 mg/l of NO_3-N present in the influent.

The biofilm contains the major portion of active biomass and thus accounts for 89 to 97% of the reactions occurring in the system. The tank reactions are seen to account for 3.4 to 10.8% of the total reaction, the higher values for the higher suspended solids concentrations carried in the tanks. The liquid film removals are negligible due to the relatively low volume compared to the tank volume. In viewing the nitrogen removal data, less than 4% of the total nitrogen removed from the system occurs by stripping at pH values less than 7.7. At higher pH values, greater ammonia stripping will occur; data set 6 showing 7.1% of the nitrogen removal occurred by stripping due mainly to the stage 1 pH of 8.7. The remainder of the nitrogen removal occurs by net growth and denitrification.

At lower temperatures, growth predominates due to the higher dissolved oxygen values in the tank and lower heterotrophic endogenous decay rates; both reducing denitrification.

Applications of the RBC Model

After obtaining kinetic rate constants from pilot plant data, the RBC model can be employed to design full-scale plants to minimize total treatment cost. A significant component of total treatment cost is the capital expenditure for media surface. It is therefore advantageous to obtain a high BOD removal per unit of media surface area for each treatment stage.

As illustrated in Figure 14, the RBC model can be employed to find optimum operating conditions for carbonaceous removal. Removal increases with BOD applied up to a maximum value where oxygen limitation results. Consider the treatment of a wastewater containing 100 mg BOD/1 and zero dissolved oxygen. From Figure 14, it is seen that a maximum BOD removal is achieved by sizing the first treatment stage so that the applied BOD rate is 7.5 lb/d/1000 sf (36.7 kg/d/1000 sq. m). Under these conditions, the BOD removal rate will be 3.9 lb/d/1000 sf (19.1 kg/d/1000 sq. m) corresponding to 52 percent removal. Thus, the influent to the second stage will be a substrate concentration of 48 mg BOD/1, and a similar procedure can be utilized. This leads to a "T" type design configuration, a large first stage followed by smaller later stages. If the design incorporates applied organic loadings much greater than the breakpoints in Figure 14, excess biological growth should result on the media since an excess quantity of substrate is available to provide anaerobic growth in the lower biofilm layers.

In a recent paper, Chesner and Molof[9] observed that BOD removals decrease with increasing media diameter of a laboratory RBC system operated at constant peripheral velocity. This is a very significant finding, in view of the fact that full-scale plants are often designed by scaling-up pilot plant data at fixed peripheral speed and hydraulic loading.

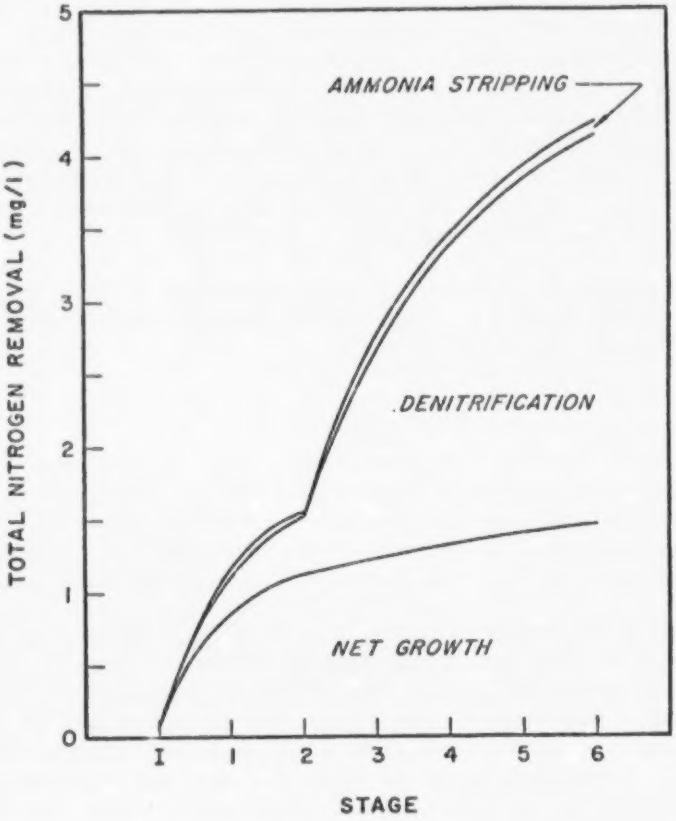

FIGURE 13
DISTRIBUTION OF TOTAL SOLUBLE NITROGEN REMOVAL
FOR DATA SET I

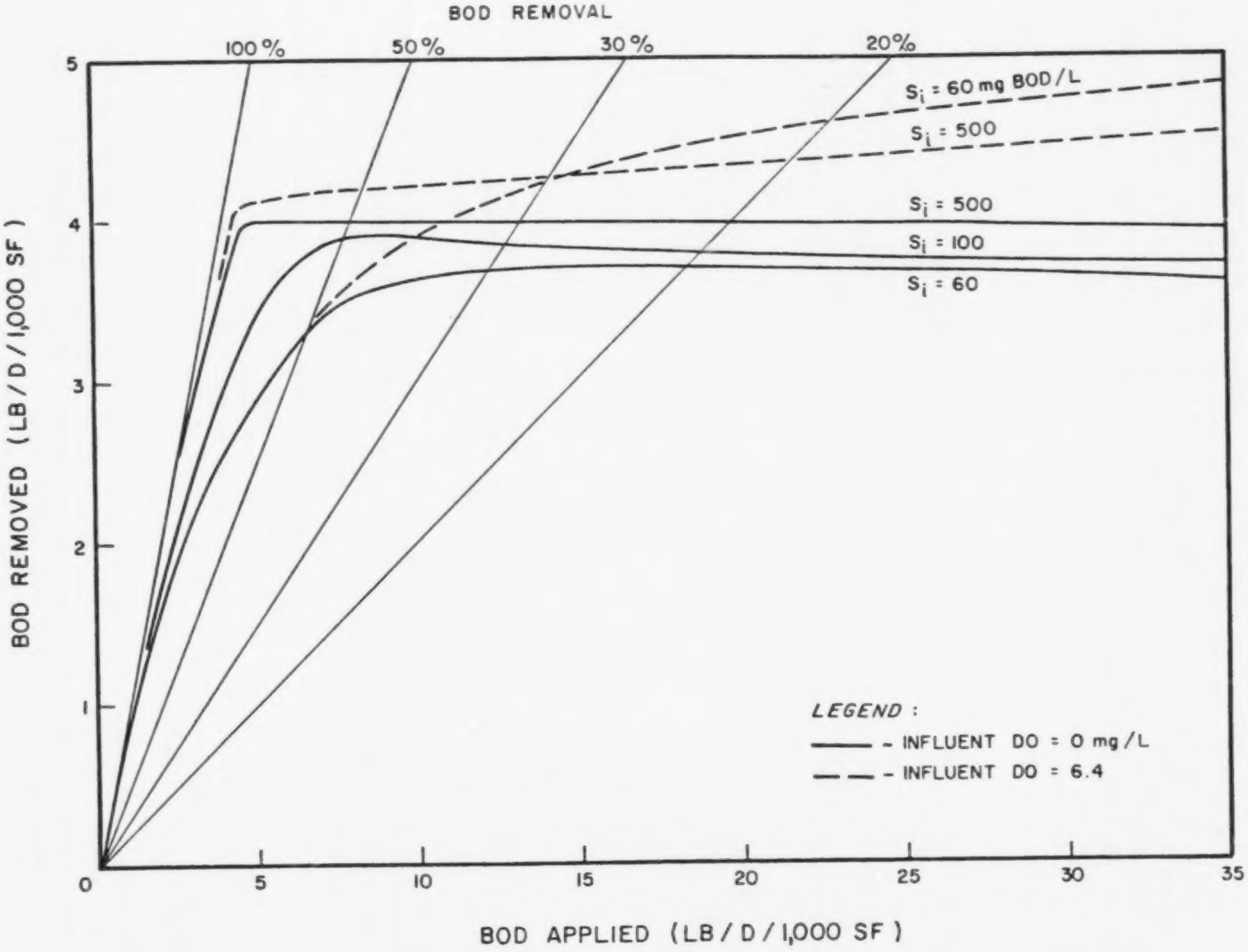

FIGURE 14
SINGLE STAGE BOD REMOVALS PER MEDIA SURFACE VS. APPLIED BOD

1 lb/d /1000 sf = 4.89 kg./d /sq. m

The RBC model predicts this effect for the 6 gpd/ft^2 pilot plant run given in Figure 4. In order to maintain constant peripheral velocity, rotational velocity must be decreased as disc size is increased. Detention time of the liquid film above the water line is increased, resulting in a large reduction in substrate concentration before reentry of the media into the mixed liquid in the tank. For a peripheral velocity of 1 fps, the detention times above the water line are 10.3 and 31.4 seconds for the 6.56 foot (2 m) and 20 foot (6.1 m) discs, respectively. In stage 1, substrate concentration dropped to 15.4 mg BOD/1 on the 6.56 foot pilot unit, and to 2.2 mg BOD/1 on the 20 foot disc. Low concentrations in the liquid film coating the biofilm result in a depletion of substrate within the biofilm. As a consequence, large portions of the biofilm are inactive above the water line for the 20-foot media.

The effect on total BOD removal for the four-stage system (Figure 15) is a reduction from 86 to 77.5 percent in going from the 6.56 foot pilot disc to a 20 foot plant disc. Pitfalls of this type can be avoided by designing plants with the RBC model after coefficients have been obtained from pilot scale data. Otherwise appropriate scaling factors have to be obtained from comparative studies using different diameter systems on the same wastewater (Wilson et al[10]).

Afer having sufficiently verified the above approach with pilot and full scale data, the model provides the framework necessary to analyze additional operating modes and conditions, thus minimizing the relatively expensive and time consuming pilot or full scale studies. Utilizing the highest loading condition for the nitrifying six stage Gladstone, Michigan plant (Figure 7), Figure 16 shows the effect of providing a high purity oxygen atmosphere in the aerated sectors of the RBC system. The higher driving force for oxygen transfer to the aerated sectors reduces the oxygen limitation on the nitrification rate, allowing a lesser number of stages to provide the same effluent quality. No effect on BOD removal results since the low BOD concentrations cause the heterotrophs to be substrate limited rather than oxygen limited.

Utilizing air instead of mechanical drive to rotate the media provides an additional source of oxygen transfer to the system by diffused aeration in the mixed tank. The oxygen absorption efficiency is probably somewhat lower than coarse bubble aeration due to the lower liquid depths. As shown in Figure 17, use of an air drive system with absorption efficiencies of either 2% or 4% will provide higher tank dissolved oxygen concentrations, and will require fewer stages than a mechanical drive system to achieve similar effluent qualities. The air drive system at 4% oxygen absorption efficiency would provide efficiencies similar to those using a 30% oxygen atmosphere in the gas phase (C_s = 15 mg/l).

The effect of temperature on nitrification efficiency is shown in Figure 18 as a function of hydraulic loading rate for a given influent composition and system design, the lower temperature causing higher effluent TKN values at similar loading rates. Optimum utilization of RBC systems will result if a variable effluent standard for TKN was allowed, higher in the winter than in the summer based on receiving water quality considerations.

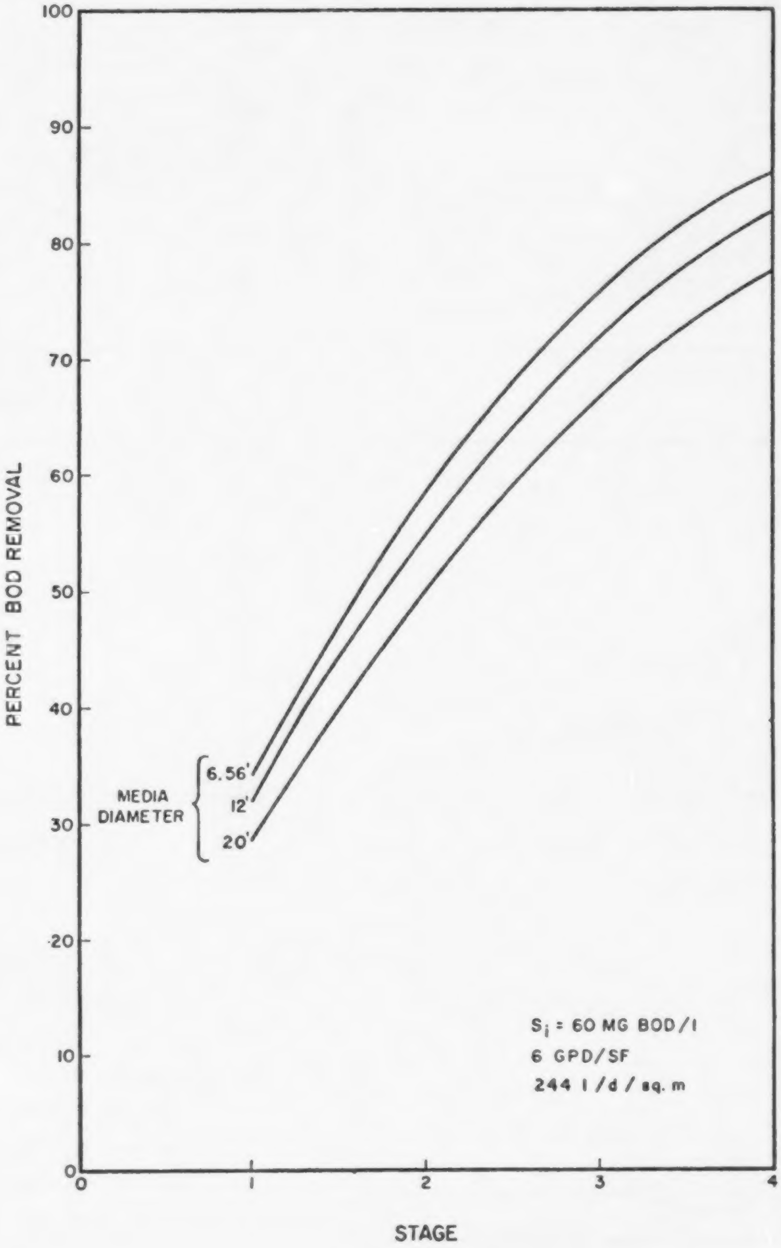

FIGURE 15
EFFECT OF MEDIA DIAMETER ON
BOD REMOVALS AT CONSTANT PERIPHERAL VELOCITY

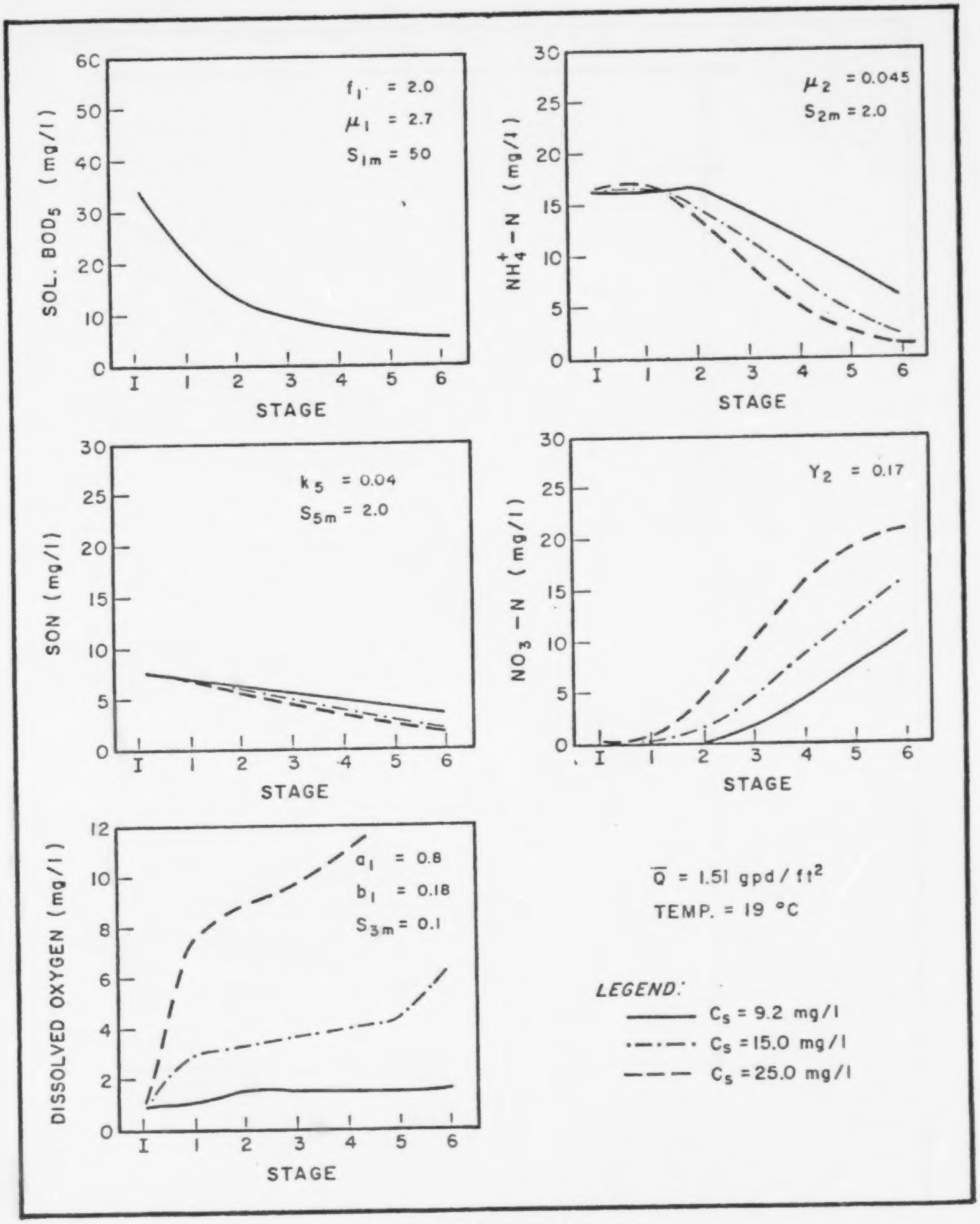

FIGURE 16

EFFECT OF OXYGEN SATURATION ON RBC PERFORMANCE

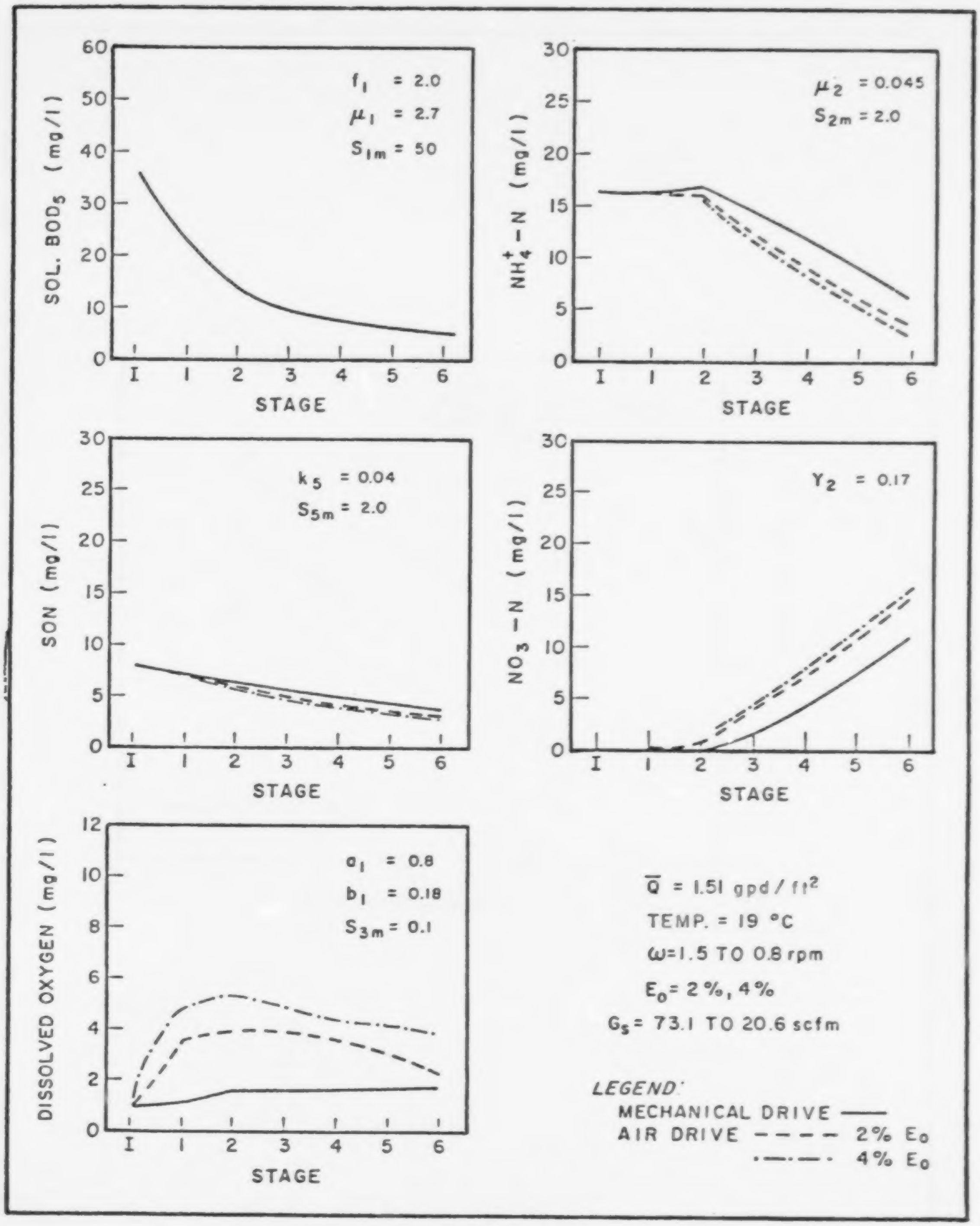

FIGURE 17
MECHANICAL DRIVE VERSUS AIR DRIVE (TAPERED AERATION)

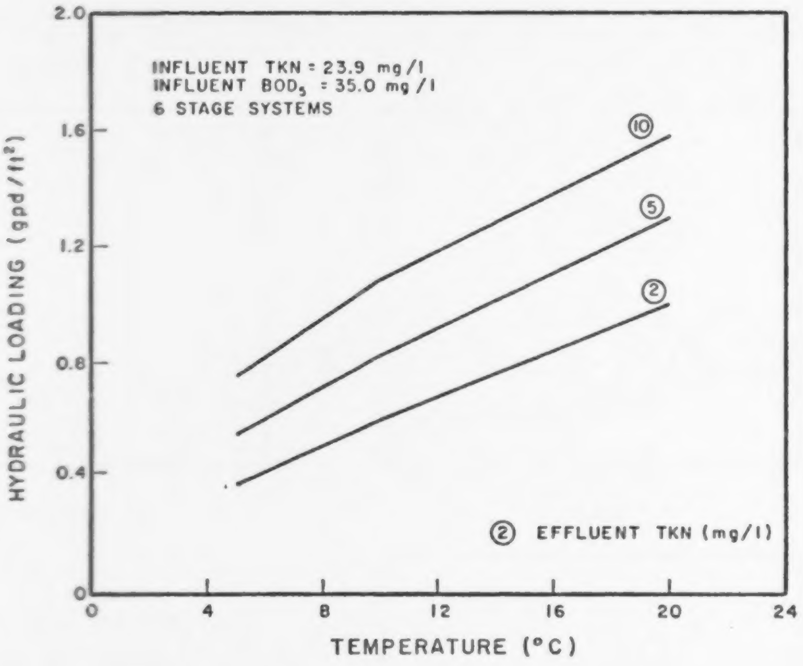

FIGURE 18

HYDRAULIC LOADING VERSUS TEMPERATURE
FOR VARIOUS EFFLUENT TKN REQUIREMENTS

Figure 19 evaluates the effect of typical municipal hourly load and flow fluctuations on RBC system performance. The biological growth distribution between autotrophs and heterotrophs was held constant at the distribution obtained for steady state conditions. The load fluctuations over a day for a given system design will cause significant fluctuations in effluent quality over the day, the peak flows and loads causing increased effluent concentrations, while nighttime low flow values would have significantly lower concentrations. This agrees well with the data of Filion et al[11].

CONCLUSIONS

1. Organic carbon removal, nitrification, and total nitrogen removal in the RBC process can be successfully described by incorporating the basic principles of mass transfer and simultaneous reaction in a steady state mathematical model.

2. Design of a full scale plant from pilot plant data without use of the model or appropriate scale-up factors can result in underdesign.

3. From the modeling results and verification studies conducted to date, it is concluded that:

   a) The degree of ammonia stripping in the RBC process is slight (normally less than 4% of total nitrogen removed). This is due to the relatively low gas exchange rates and low non-ionized fractions at pH values less than 8.0.

   b) The major fraction of total nitrogen removal occurs due to growth and denitrification, the former predominating during low temperatures due to high D.O. values. The major portion of denitrification occurs in the middle stages where nitrification is initiated in the presence of significant heterotrophs.

   c) Reactions occurring in the liquid film in the aerated sectors are negligible due to low liquid volume and suspended solids concentrations.

   d) Tank reactions are a function of the tank MLVSS concentrations in each stage. Less than 5% of total reaction occurred in the tank at MLVSS concentrations of 50-80 mg/l while as much as 11% occurred at concentration above 110 mg/l.

4. Calibration of the model for air drive systems is required to determine appropriate O_2 transfer efficiencies and effects on kinetic coefficients.

5. After calibration, the model allows the designer to describe process performance with one set of temperature-corrected kinetic coefficients for a particular waste. It puts the description of the process on a rational basis which will be useful in analyzing results of full and pilot scale studies. It should provide the basis for design since it allows evaluation of various process configurations and wastewater characteristics other than those of the pilot study.

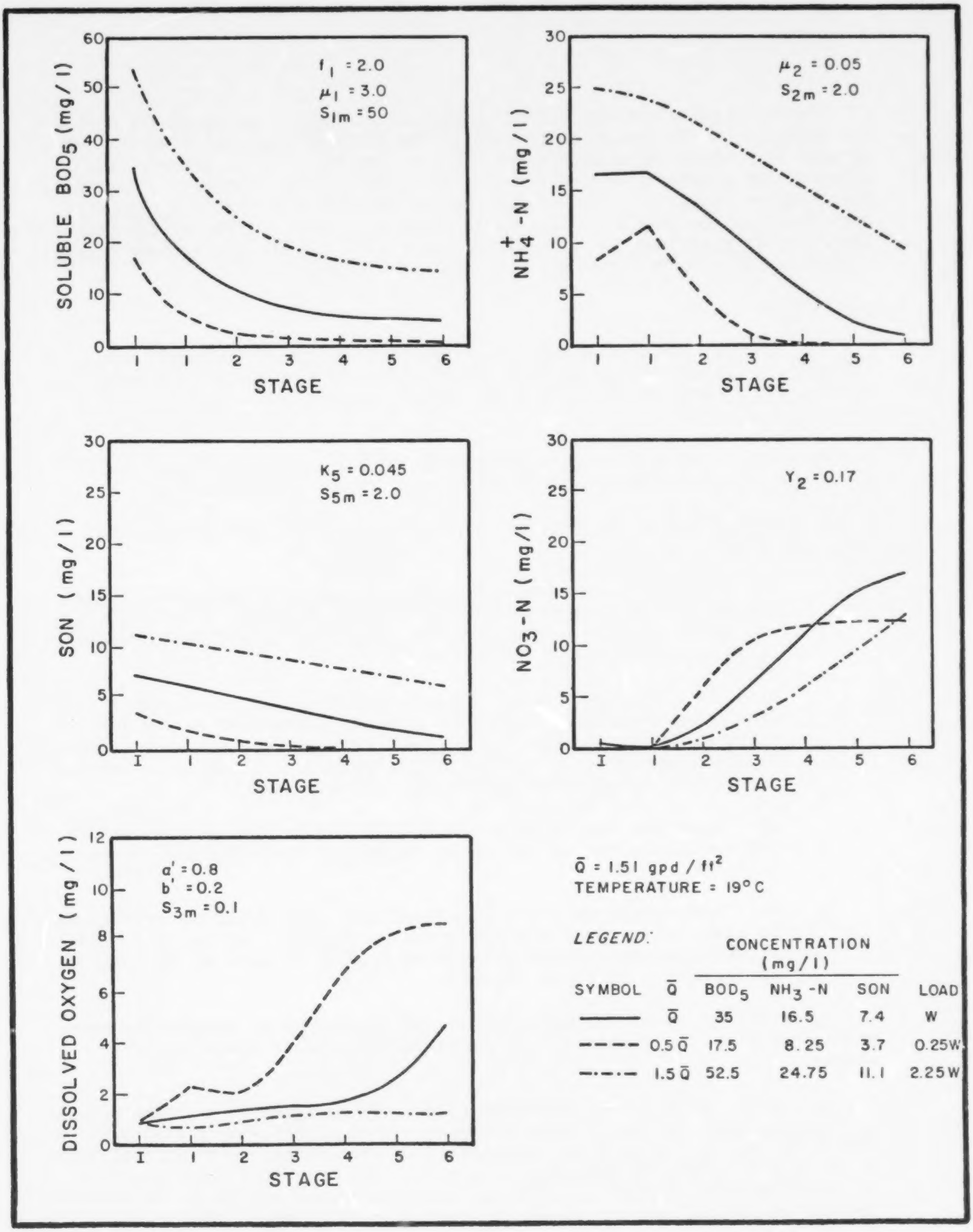

FIGURE 19
EFFECT OF LOAD VARIABILITY ON RBC PERFORMANCE

NOMENCLATURE

Parameters	Subscript	Superscript	Identification	Dimensions
A			Interfacial area	L^2
	T		Total wetted Area for one side of clean disc	
	A		Total aerated area	
D			Diffusivity	L^2/T
	S		Component diffusivity in biofilm	
	S	L	Component diffusivity in liquid film	
E	o		O_2 absorption efficiency for air drive	%
EX			Endogenous respiration rate	M/L^3-T
G			Gas flow rate	L^3/T
	o		Air flow rate for air drive media	scfm
H	2		Henry's constant for ammonia	mg/l-atm
K	L		Transfer coefficient from gas to liquid	L/T
	S		Transfer coefficient from liquid to biofilm	L/T
K_La			O_2 transfer coefficient in tank for air drive	T^{-1}
k	5		SON reaction rate	T^{-1}
M	2		Molecular weight ammonia	g/mole
N	W		Number submerged sectors	
Q			Influent wastewater flow rate	L^3/T
	L		Liquid film flow rate	
	F		Biofilm flow rate	

NOMENCLATURE (Continued)

Parameters	Subscript	Superscript	Identification	Dimensions
R			Reaction rate	M/L^3-T
R			Gas constant	0.082 l-atm/g-mole-°K
	H		Hydraulic radius of channels	L
RX			Growth rate	M/L^3-T
S			Component concentration in Biofilm	
	1		Substrate (BOD_5 or COD)	
	2		Ammonia Nitrogen	
	2N		Nonionized ammonia nitrogen	
	3		Oxygen	
	4		Nitrate nitrogen	
	5		Soluble organic nitrogen	
		L	Liquid film	
		T	Tank	
		0	Influent	
		1	1st biofilm layer	
		L4	Last aerated sector	
T			Absolute Temperature	°K
t			Temperature	°C
V			Tank volume	L^3
	g	-	Gas volume adjacent to wetted area, A_T	
X			Organism (VSS) concentration	mg/l
	1		Heterotrophs	
	2		Autotrophs	

NOMENCLATURE (Continued)

Parameters	Subscript	Superscript	Identification	Dimensions
δ			Active biofilm thickness	L
	L		Liquid film thickness	
ω			Rotational speed	rpm

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PART VIII: UPGRADING PRIMARY AND SECONDARY WASTE TREATMENT SYSTEMS WITH RBC

UPGRADING PRIMARY TANKS WITH ROTATING BIOLOGICAL CONTACTORS

by

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Introduction

The Borough of Edgewater, New Jersey, operates a primary wastewater treatment facility which discharges into the Hudson River. The State of New Jersey, in 1971, imposed a requirement on Edgewater to upgrade its facility to secondary treatment. Constrained by severe land limitations, several treatment alternatives were considered which would minimize plant expansion.

A process which indicated considerable promise involved the installation of rotating biological contactors (RBC) in the existing primary clarifiers. An intermediate floor would be installed at mid-depth in the tanks. Biological treatment of the raw wastewaters would be accomplished in the upper RBC sector and secondary clarification would take place in the sector below the floor. Because the proposed treatment scheme involved new concepts, a pilot study was required to confirm the feasibility of the process and to develop design and cost information. A program was then developed and financed by Edgewater to evaluate the RBC/Underflow Clarifier system with a prototype unit.

The installation of the RBC/Underflow Clarifier pilot facility was completed in May 1973. The process evaluation was conducted over a period of three years by Edgewater personnel and results from these studies indicated that modification of the primary clarifier to the two-tier treatment process could produce a secondary treatment effluent commensurate with U.S. Environmental Protection Agency effluent requirements. Realizing its potential, Edgewater officials sought, and received, a demonstration grant from the U.S. EPA to improve the existing facility and to continue the process evaluation under an intensive testing program.

OBJECTIVES AND SCOPE OF STUDY

The objectives of the RBC/Underflow Clarifier pilot evaluation were as follows:

(1) Establish the feasibility of upgrading existing primary sedimentation plants to meet the secondary treatment requirements of PL 92-500 through the installation of RBC units directly in primary clarifiers.

(2) Evaluate the degree of pretreatment necessary to successfully operate an RBC system in this mode.

(3) Evaluate the effects of climatic conditions, diurnal flow, and total daily load and waste characteristic variations on process efficiency.

(4) Establish process and plant design parameters, and capital and operating costs for the application of this upgrading technique to maximize the use of tankage and facilities at existing primary sedimentation plants.

This paper attempts to highlight several aspects of the study. A formal report was prepared and submitted to the U.S.E.P.A. Municipal Environmental Research Laboratory, Cincinnati, Ohio, which documents the results of the total program.

EDGEWATER TREATMENT PLANT

The Edgewater Sewage Treatment plant has a design capacity of 1140 m^3/d (3 mgd), and provides primary treatment of combined sanitary and storm waters. Figure 1 shows a plan of the existing plant site. The major facilities include an administration building, pump house, comminutor, grit collector, five primary settling tanks, chlorine building, and outfall sewer. Sludge is processed in two anaerobic digestors and two vacuum filters. A flash dryer is available although not presently used. Land is limited, comprising only 1.5 acres of usable area.

Table 1 presents a characterization of the raw wastewater for the one year testing period from March 1977 to February 1978.

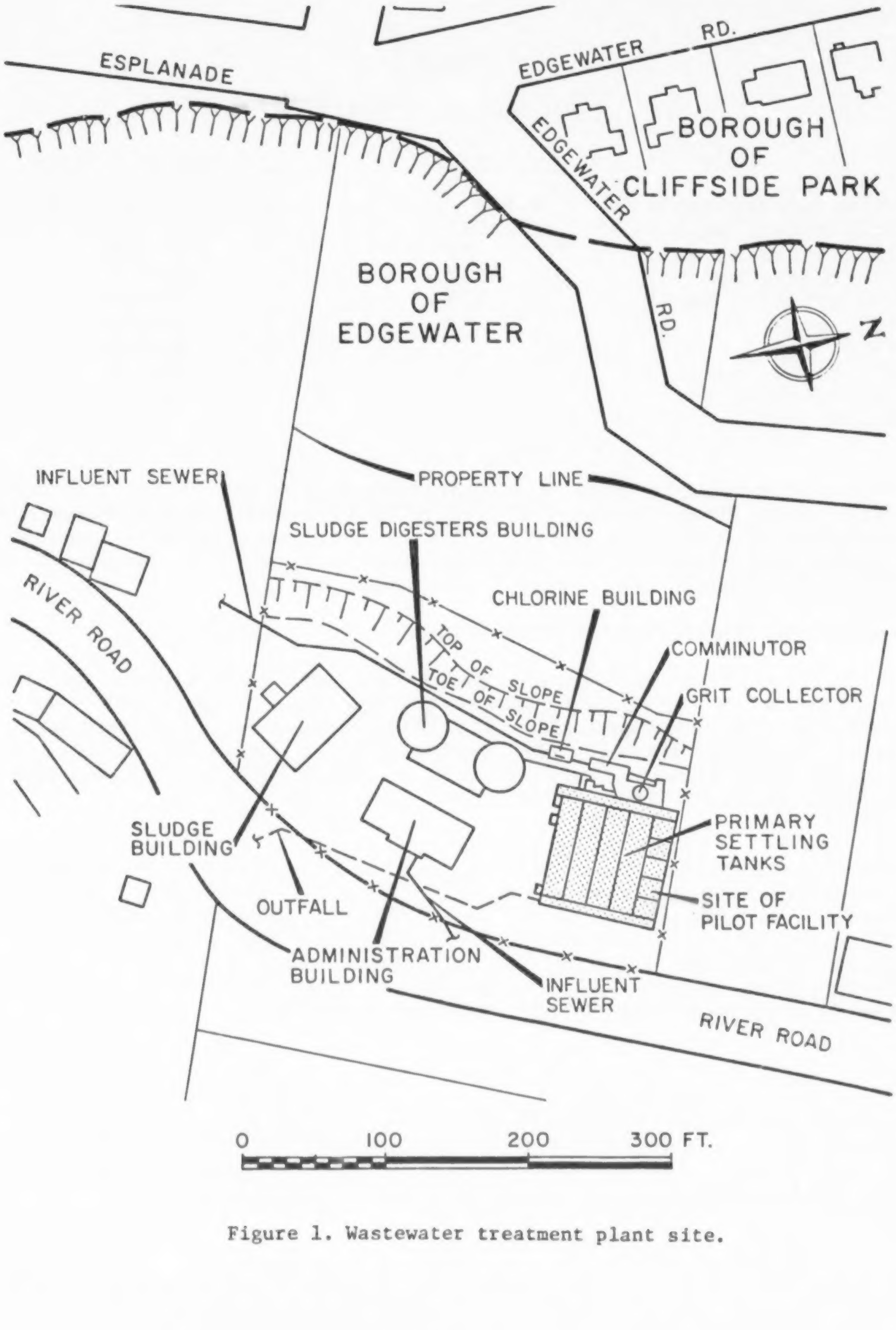

ESPLANADE
EDGEWATER RD.
EDGEWATER RD.
BOROUGH OF CLIFFSIDE PARK
BOROUGH OF EDGEWATER
N
PROPERTY LINE
INFLUENT SEWER
SLUDGE DIGESTERS BUILDING
CHLORINE BUILDING
COMMINUTOR
GRIT COLLECTOR
RIVER ROAD
TOP OF SLOPE
TOE OF SLOPE
PRIMARY SETTLING TANKS
SITE OF PILOT FACILITY
SLUDGE BUILDING
OUTFALL
ADMINISTRATION BUILDING
INFLUENT SEWER
RIVER ROAD
0
100
200
300 FT.

Figure 1. Wastewater treatment plant site.

TABLE 1

RAW SEWAGE COMPOSITION

	Average	Range of Values
Flow, m^3/day (mgd)	9,920 (2.6)	4,540-31,800 (1.2-8.4)
BOD_5 total, mg/l	144	50-573
BOD_5 soluble, mg/l	80	22-188
COD total, mg/l	350	128-772
COD soluble, mg/l	176	67-280
TSS, mg/l	169	36-280
TVSS, mg/l	137	44-206
TKN total, mg/l	26	10- 41
TKN soluble, mg/l	22	9- 31
NH_3-N, mg/l	13	3- 21

The average daily flow was approximately 9,900 m^3/d (2.6 mgd). Peak storm flows exceeded 27,000 m^3/d (7.2 mgd), the maximum flow capacity of the flow meter. The raw wastewaters were generally representative of dilute combined sewerage, with an average total BOD_5 of 144 mg/l and 169 mg/l suspended solids.

RBC/UNDERFLOW CLARIFIER TEST MODULE

A layout of the primary clarifiers and test module is shown on Figure 2. Primary Settling Tank No. 5 was converted to evaluate the RBC/Underflow clarifier concept. Raw wastewater, after passing through the comminutor and grit collector, was pumped from a point located 3.51 m (11.5 ft) from the influent side of Settling Tank No. 3.

Figure 3 shows a cross-section of Tank 5 after conversion. The intermediate floor was installed to provide an underflow clarifier with a water depth of 1.42 m (4 ft 8 in). Four RBC units with diameters of 3.61 m (12 ft) were installed in the 21.34 m (70 ft) long by 4.37 m (14 ft 4 in) wide tank with a water depth of 1.22 m (4 ft) above the intermediate floor. Fiberglass covers were installed over the RBC units for protection of the media and attached biomass.

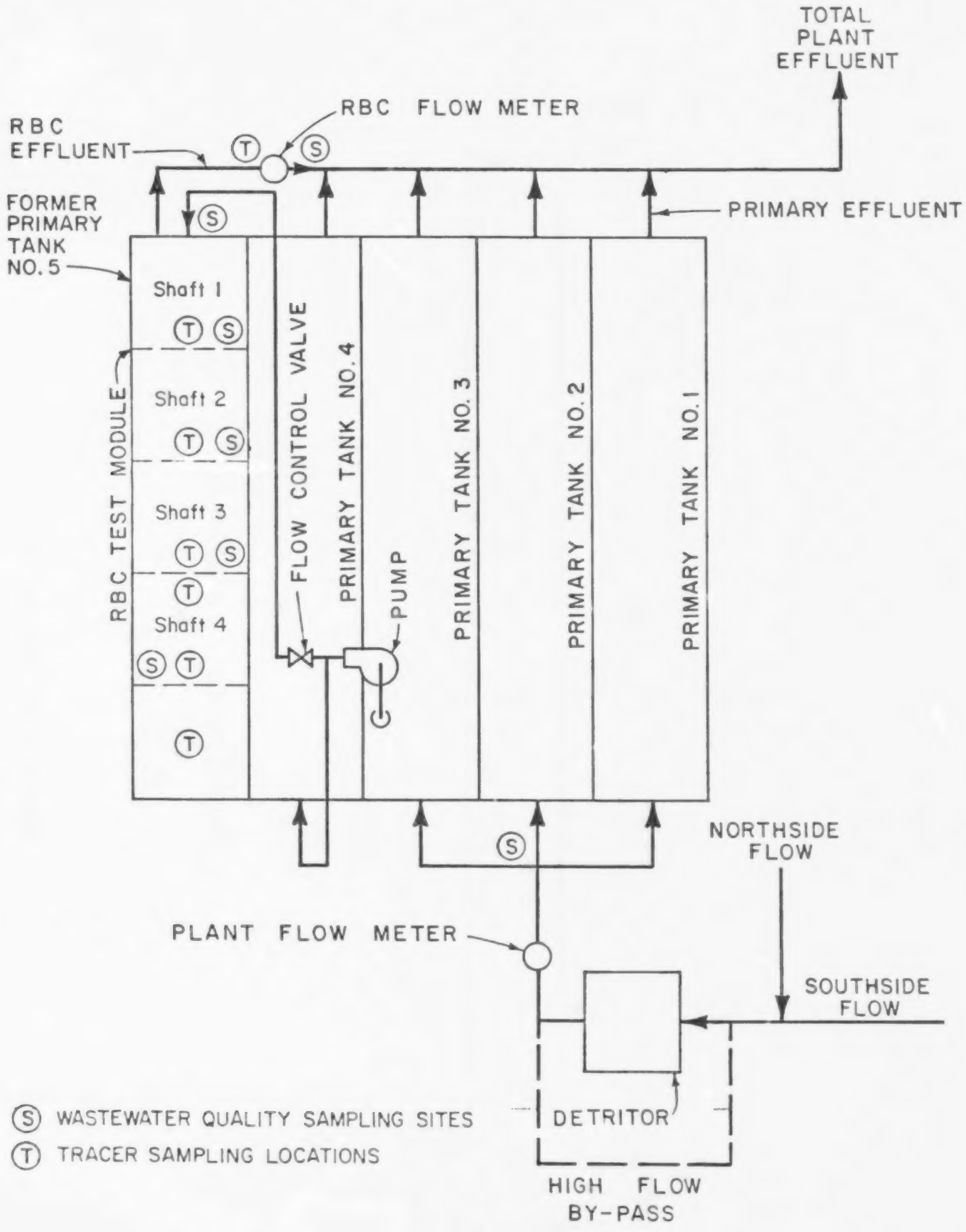

TOTAL PLANT EFFLUENT
RBC FLOW METER
RBC EFFLUENT
Ⓣ
Ⓢ
FORMER PRIMARY TANK NO. 5
PRIMARY EFFLUENT
Shaft 1
Shaft 2
Shaft 3
Shaft 4
RBC TEST MODULE
FLOW CONTROL VALVE
PRIMARY TANK NO. 4
PUMP
PRIMARY TANK NO. 3
PRIMARY TANK NO. 2
PRIMARY TANK NO. 1
NORTHSIDE FLOW
PLANT FLOW METER
SOUTHSIDE FLOW
DETRITOR
HIGH FLOW BY-PASS
Ⓢ WASTEWATER QUALITY SAMPLING SITES
Ⓣ TRACER SAMPLING LOCATIONS

Figure 2. Plant flow schematic.

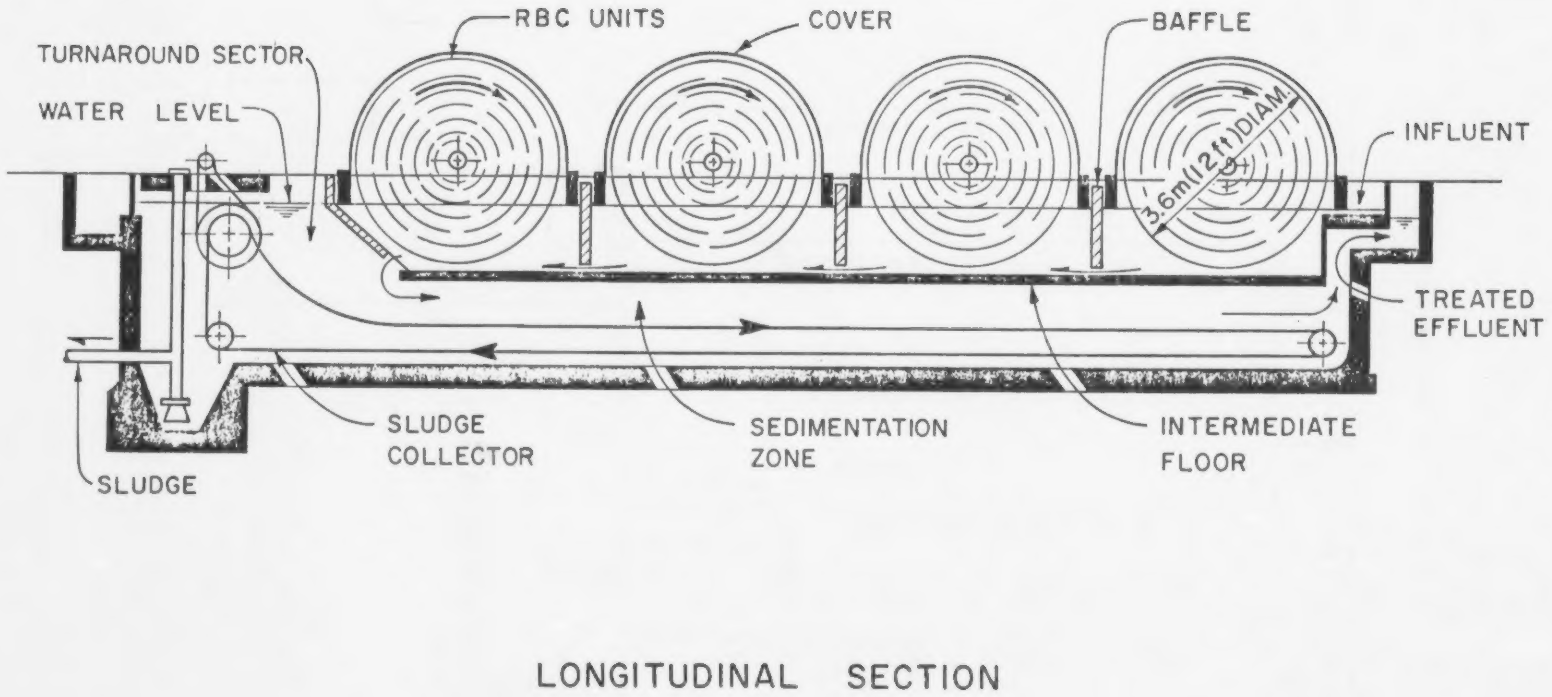

RBC UNITS
COVER
BAFFLE
TURNAROUND SECTOR
WATER LEVEL
INFLUENT
3.6m(12ft)DIAM.
TREATED EFFLUENT
SLUDGE
SLUDGE COLLECTOR
SEDIMENTATION ZONE
INTERMEDIATE FLOOR
LONGITUDINAL SECTION

Figure 3. RBC testing unit.

The RBC units were made of high-density polyethlyene. Stages one, two and three each had a surface area of 1,220 m^2/m of shaft length (4,000 ft^2/ft) and stage four had a surface area of 1,830 m^2/m (6,000 ft^2/ft). Each of the four shafts was 4.1 m (13 ft 5 in) long. The units were immersed to a depth of 1.07 m (3 ft 6 in) which provided a total effective wetted surface area for the four shafts of 18,200 m^2 (196,500 ft^2). The small portion of the central surface which was not wetted and effectively free of microorganisms represented 17 percent of the total surface area. Throughout this paper, loading will be expressed on an effective surface area basis, i.e., wetted surface area.

The test program anticipated the removal of grit and trash without removing substantial portions of BOD and SS. Since the plant detritor could be by-passed at times of high influent flows to the plant, a modification was introduced to provide the intake to the RBC pump at a point 3.51 m (11 ft 6 in) from the head end of Settling Tank No. 3. This allowed high rate primary treatment prior to the RBC system.

A pump capable of providing a maximum test flow of 5,680 m^3/day (1.5 mgd) was installed, with a programmer to provide a variable signal to an electrically operated valve. The valve was programmed to provide a diurnal hydraulic variation, with the peak to average ratio equal to 1.5 and the minimum to average ratio equal to 0.5.

A weir was installed in the influent channel to the RBC sector to distribute the flow uniformly. Additionally, screens were attached to the influent channel to catch large fibrous material. The intermediate floor and the influent channel were adequately caulked to prevent leakage and/or exchange of effluent and incoming wastewaters.

EXPERIMENTAL PROGRAM

Phase 1 of the experimental program studied the RBC system performance over a wide range of loading conditions. The initial loading was set relatively low to ensure an effluent quality greater than EPA requirements. (EPA secondary treatment standards presently call for 85 percent BOD_5 and TSS removal or monthly average BOD_5 and SS concentrations less than or equal to 30 mg/l. Maximum weekly average BOD_5 and SS concentrations must be less than or equal to 45 mg/l.) The loading was then increased to yield an effluent quality approximately equivalent to EPA standards. The third and final loading condition was chosen to stress the RBC system, i.e. violate the 30/30 BOD_5/SS standards.

The low loading condition was evaluated for approximately two weeks. The moderate and high loading conditions were each evaluated over an approximate period of five weeks. Several days were provided before each analysis period for the system to acclimate to the change in loading. Typically, this acclimation period extended over one to two weeks.

Phase 2 involved selection of an optimum system loading rate based on an analysis of the data collected from Phase 1. This selection was aided by use of a computer simulation model of fixed film kinetics with particular application to the RBC system. The second phase of the program studied long-term steady state operation of the RBC at the pre-selected optimum loading rate applied during warm temperature conditions and low dissolved oxygen levels. A two-week acclimation period was provided before this study period, which lasted two months.

Phase 3 of the experimental program imposed the optimum loading on the system during winter, cold temperature conditions for a period of 2-1/2 months. The loading was maintained at or near that evaluated during the summer months.

Sampling

Seven sampling locations were utilized to monitor water quality through the RBC system. These are shown on Figure 2. Daily, 24-hour, flow-proportioned composite samples were drawn from the raw influent, RBC influent, and the final effluent from the RBC/Underflow Clarifier. Discrete samplers, ISCO Model Number 1680 Samplers with Multiplexers, Model Number 1295, were positioned at the RBC influent and underflow clarifier effluent; a single composite sampler, ISCO Model Number 1580W, was maintained in the raw influent waste stream.

Periodically, 24-hour flow-proportioned composite samples were drawn from each of the four stages in the RBC system. These samples were drawn from mid-depth with submersible pumps, and composited in 18.9 l (5 gal) containers packed with ice and insulated.

Analytical Program

Table 2 summarizes the analytical schedule followed during the major phases of the experimental program. The numbers indicate the number of samples analyzed for that particular parameter per week.

Studies were also conducted periodically to characterize the physical and hydraulic operation of the system. These included tracer analyses, zone and flocculant settling tests, and chemical addition tests.

During each flow condition, or major modification to the physical system, a tracer analysis was conducted to characterize the hydraulics through the RBC system and to monitor the system for any physical abnormality such as leakage, etc. Lithium chloride was evenly distributed across the RBC influent channel and samples taken with time at selected sampling locations (see Figure 2). The samples were then analyzed for lithium by standard atomic absorption spectrophotometer procedures.

Flocculant settling tests were conducted using 2.13 m (7 ft) high, 15.24 cm (6 in) diameter columns with sampling ports at 0.305 m (1 ft) intervals. Sample (typically from the fourth stage) was pumped into the column and aliquots drawn at each port at regular time intervals. Standard jar test procedures were employed to evaluate the effects and feasibility of chemical addition to improve solids capture in the RBC/Underflow Clarifier.

EXPERIMENTAL RESULTS

The results of the three phase program are highlighted in the following discussions.

Variable Loading

The results of the system evaluation at several loading conditions are summarized in Table 3. The flows to the RBC/Underflow Clarifier system ranged from 1000 to 2500 m^3/d (0.23 to 0.67 mgd). The total BOD_5 loads to the system were 5.3, 11.7 and 19.7 g $BOD/d/m^2$ (1.1, 2.4 and 4.04 lb $BOD_5/d/1000\ ft^2$) for the low, moderate and high loading conditions, respectively.

TABLE 2

ANALYTICAL SCHEDULE
(NUMBER OF SAMPLES ANALYZED PER WEEK)

Analysis(4)	Raw Influent	RBC Influent	Stages 1,2,3&4	RBC Effluent	RBC Sludge
Flow	Recorded			Recorded	When Drawn
Temperature		7		7	
pH	7	7	7	7	
DO	7	7	7	7	
BOD$_5$ (T)	7	7		7	
BOD$_5$ (S)	5	5	2(1)	5	
COD (T)	5	5		5	
COD (S)	5	4	2(1)	5	
TS					When Drawn
TVS					When Drawn
TSS	7	7	2(1)	7	
TVSS	7	7	2(1)	7	
TKN (T)(2)	3	3	1(1)	3	
TKN (S)(2)	3	3	2(1)	3	
NH$_3$-N(2)	3	3	2(1)	3	
NO$_2$-N(2)	3	3	2(1)	3	
NO$_3$-N(2)	3	3	2	3	
SO$_4$		1	1(1)	1	
Total Sulfide(2)		1	1(1)	1	
Grease/Oils (T)(3)	Biweekly	Biweekly		Biweekly	
PO$_4$-P total(3)		Periodically			
PO$_4$-P T-ortho(3)		Periodically			

(1) Only during interstage studies.
(2) Conducted at EPA Laboratories, Cincinnati, Ohio.
(3) Conducted at the Hydroscience Laboratory, Westwood, New Jersey.
(4) (T) = Total; (S) = Soluble, as defined by filtrate.

TABLE 3

EVALUATION OF LOADING VARIATIONS

	Low Loading 3/22-4/6/77	Moderate Loading 4/11-5/13/77	High Loading 5/23-6/30/77
FLow, m^3/d	1060	1440	2500
(mgd)	(0.28)	(0.39)	(0.67)
Hydraulic Loading, $m^3/d/m^2$	0.06	0.08	0.14
(gpd/ft^2)	(1.4)	(1.9)	(3.4)
Influent BOD_5, mg/l (T)	92	148	143
(S)	48	98	75
Total BOD_5 Loading $g/d/m^2$	5.3	11.7	19.7
(lbs/d/1000 ft^2)	1.1	2.4	4.0
Soluble BOD_5 Loading, $g/d/m^2$	2.8	7.7	10.4
(lbs/d/1000 ft^2)	0.6	1.6	2.1
Influent TSS, mg/l	124	122	128
Temperature, °C	13.0	17.2	23.2
Influent D.O., mg/l	6.9	5.0	2.2
Effluent BOD_5, mg/l (T)	14.5	23.2	55.0
(S)	9.7	22.1	30.7
Effluent TSS, mg/l	24	23	57

The effluent during the low and moderate loadings complied with the EPA standards for BOD_5 and suspended solids (30/30), while the stressed condition caused a violation for both BOD_5 and SS.

Interstage Analysis of the RBC System

Throughout the experimental program, 24 hour flow proportioned composite samples from each stage were analyzed on a regular basis for soluble BOD_5 and COD. Dissolved oxygen measurements were taken between 9 and 11 AM and were representative of the peak diurnal loading condition. These interstage data were used to calibrate an RBC kinetic model developed by Hydroscience. This model was in turn utilized to develop design nomographs, as discussed in a subsequent section.

The model is essentially a series of material balance equations which are solved to determine substrate and oxygen levels in the effluent from each stage and in the attached biofilm. Mass transfer resistances, determined as a function of operating conditions, are considered in both the liquid phase and biofilm, and the reaction rate is related to substrate and oxygen concentrations through the kinetic equations.

Calibration of the model was accomplished by establishing values of the variables associated with the physical and biological processes and performing a search for appropriate BOD_5 removal and oxygen utilization rate constants. Figures 4 and 5 show the final verification results for each of six study periods. The model was able to effectively predict soluble BOD_5 and DO profiles through the system using a single set of kinetic parameters for all cases. The D.O. data represent measurements taken during peak loading conditions. Model D.O. profiles are presented for both the peak and average conditions.

The analysis indicates that the model is capable of predicting system performance over a range of hydraulic and organic loading conditions using a single set of kinetic parameters. The match of the observed data demonstrates that hydraulic and mass transfer components respond correctly to system variations.

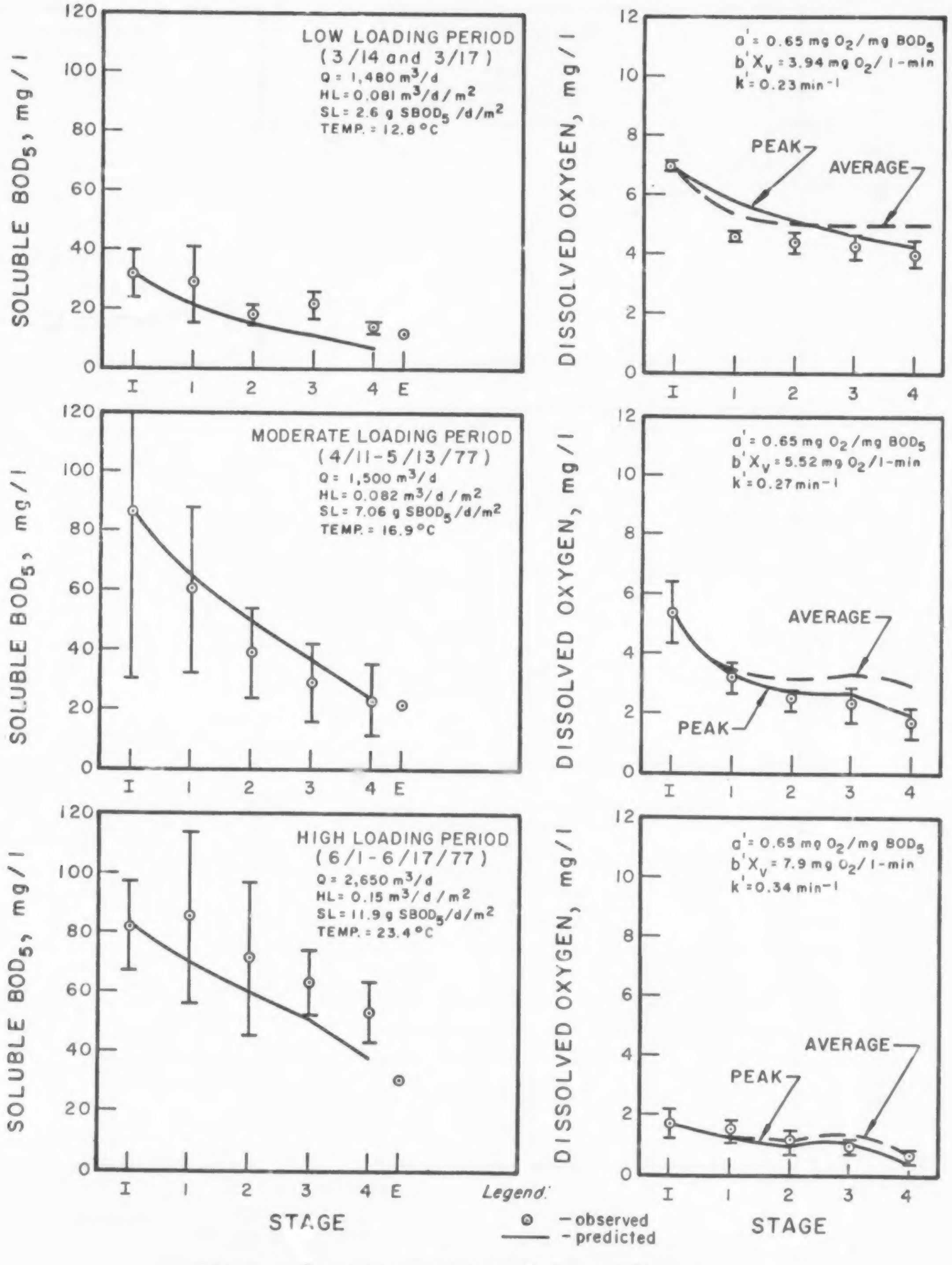


Figure 4. RBC kinetic model verification based on interstage $SBOD_5$ and D.O. data

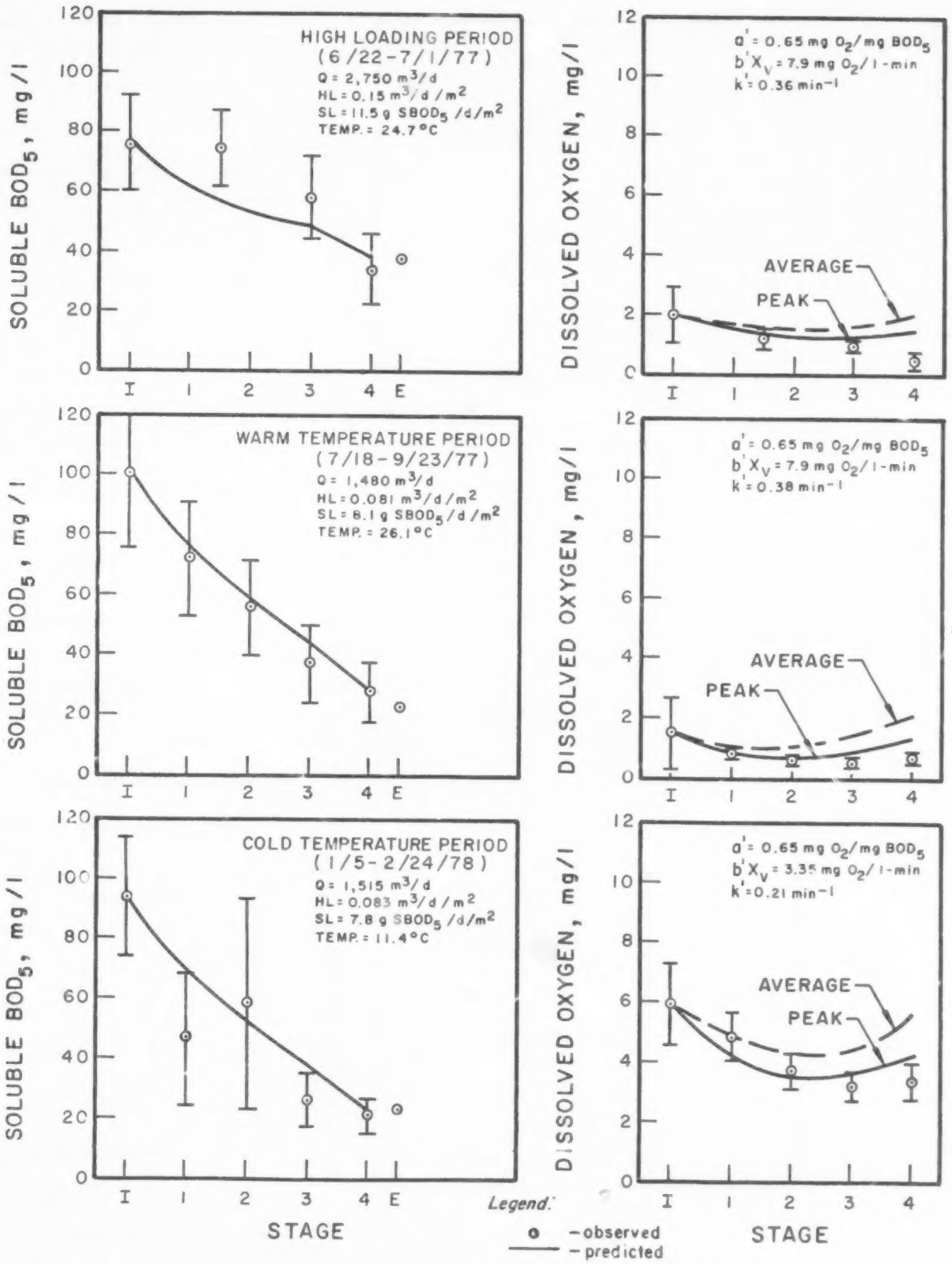


Figure 5. RBC kinetic model verification based on interstage $SBOD_5$ and D.O. data

Oxygen Utilization

A primary role of the rotating media is to provide an effective means for oxygenation of the fixed biofilm and prevent anoxic or oxygen-limiting conditions in the removal of substrate. The system can be approximated by a COD balance (assuming minimal autotrophic activity) which estimates the total oxygen utilization for both substrate oxidation and cell synthesis:

O_2 Utilization = RBC Influent COD - Effluent COD - COD Wasted

The term (influent COD - effluent COD) is effectively the TCOD removal rate and is presented on Figure 6 as a function of TCOD loading to the system. The COD equivalent of the solids was then subtracted from the TCOD removal rate and plotted as the net oxygen utilization rate, as shown on Figure 6. The oxygen utilization rate, as predicted by the kinetic model is also shown on Figure 6, and corresponds closely with the curve based on the COD balance. The flattening of the rates at the higher influent loadings suggest the system is reaching a limiting condition in its ability to transfer oxygen.

Active Biofilm

The RBC model is capable of constructing oxygen and substrate profiles throuh the RBC stages and into the biofilm. The kinetic equations will cause a reduction in substrate removal when D.O. becomes limiting. Thus, if the DO is less than an assumed Michaelis constant in regions of the biofilm, the reaction is limited by a deficiency of oxygen.

Figure 7 presents an estimation of active biofilm depth for each stage under high and moderate loading conditions.

As shown on Figure 7, the active film layer is between 300 and 600 µm, typically dictated by oxygen limiting conditions, and dependent upon loading conditions. This suggests that excessive growth of biofilm does not result in additional substrate removal. This was observed in the late summer months when acid dumps caused considerable sloughing of the attached growth. Effective treatment was still maintained with a relatively thin biofilm. A judgment as to whether the biofilm in excess of the active depth is useful is difficult. While it adds considerably to the mass to be supported by the shaft, the large solids

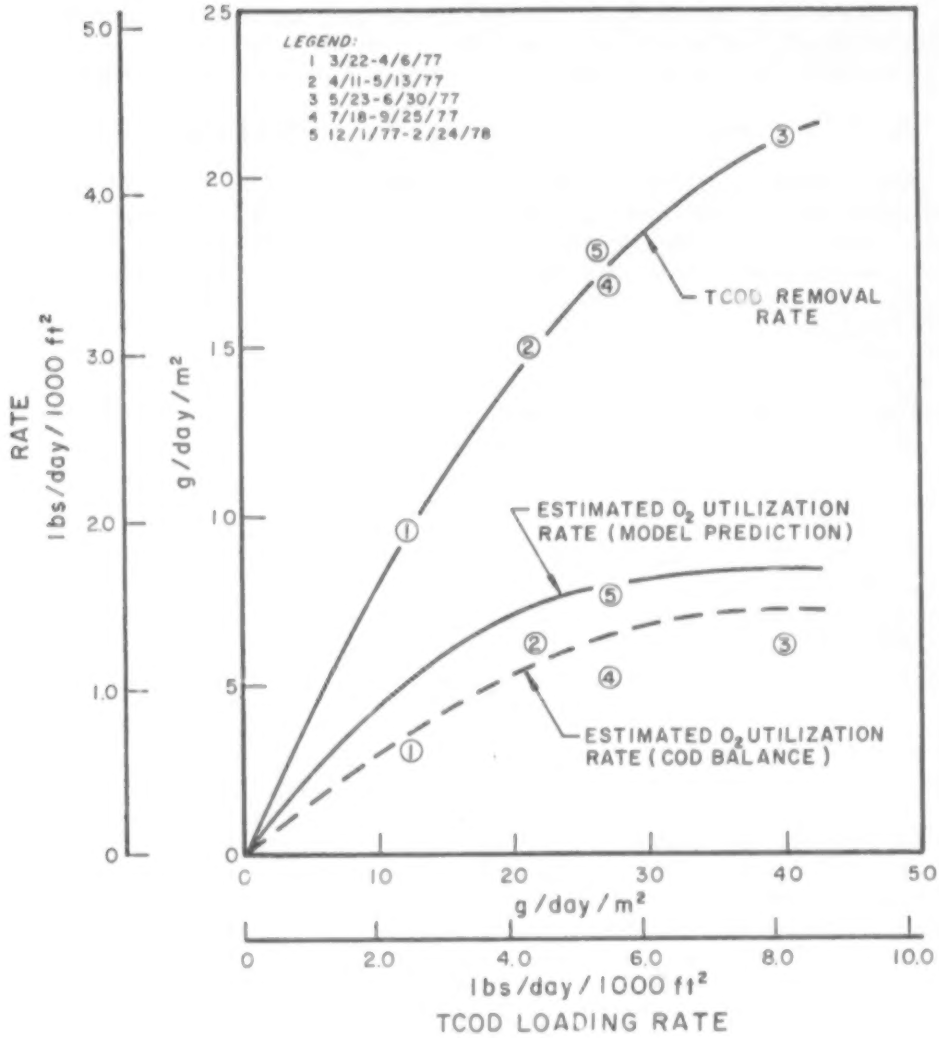
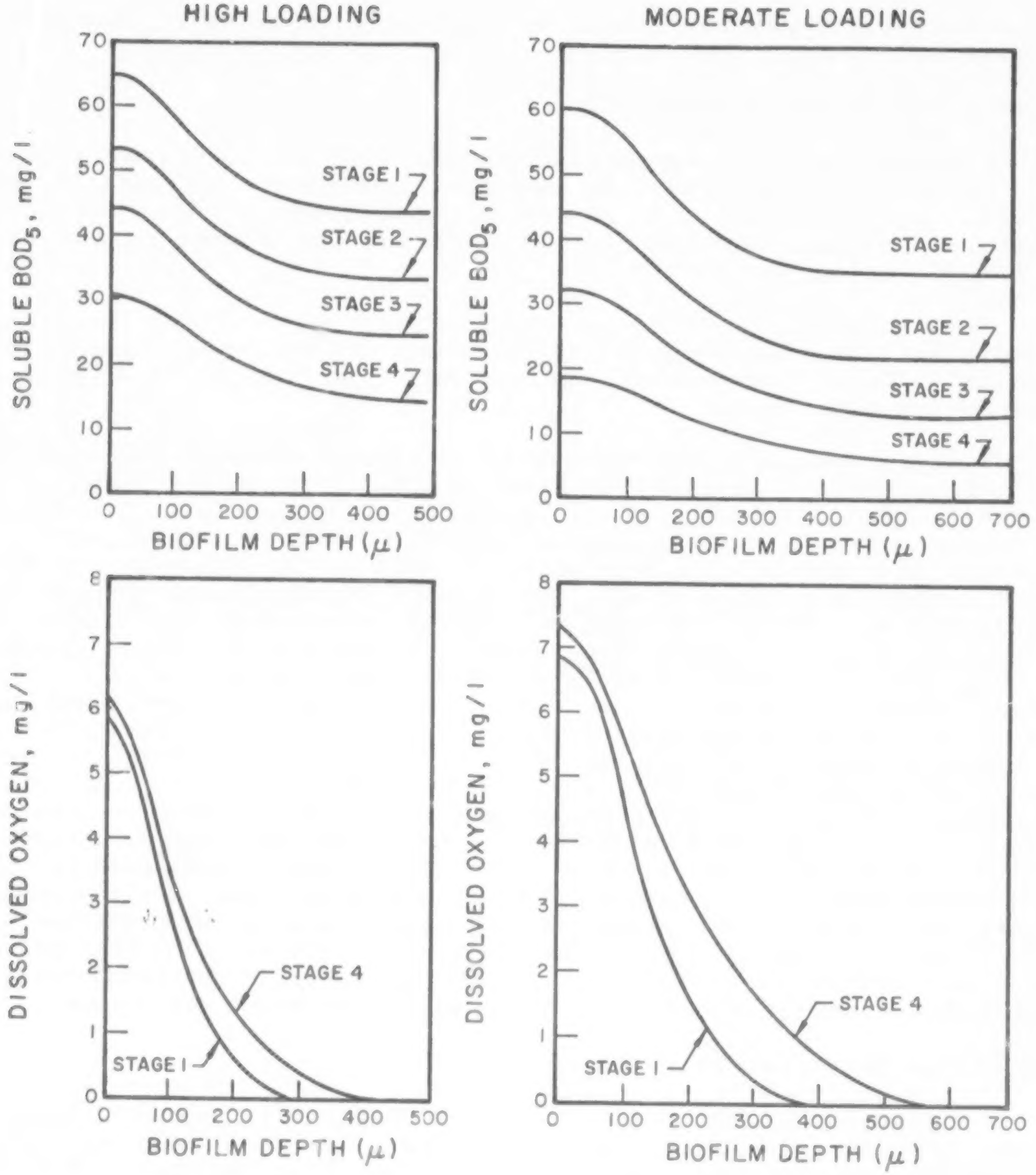


Figure 6. Estimate of RBC oxygen utilization rates



HYDRAULIC LOAD = 0.14 $m^3/d/m^2$
SOLUBLE SUBSTRATE LOAD = 10.4 $g/d/m^2$

HYDRAULIC LOAD = 0.08 $m^3/d/m^2$
SOLUBLE SUBSTRATE LOAD = 7.7 $g/d/m^2$

Figure 7. Biofilm concentrations of substrate and dissolved oxygen in successive stages

inventory may serve to control net solids production by anaerobic endogenous respiration.

Summer and Winter Evaluations

A summary of the summer and winter evaluations is presented in Table 4. There was an average temperature difference of 14.8°C between the warm and cold conditions. The influent dissolved oxygen averaged 1.4 mg/l during the summer period and 6.1 mg/l during the winter.

The results indicated that under equivalent loading conditions, similar removal efficiencies (as expressed by percent removal) were experienced during both the summer and winter evaluation periods.

Temperature will affect several of the mechanisms involved in the kinetics of the fixed film process, including substrate removal rates, oxygen saturation values (hence, mass transfer driving forces), and the diffusivities of oxygen and substrate. The minimal impact of temperature on system performance was due to compensating effects of the various parameters affected by temperature. Thus, the higher removal rates and diffusivities experienced in the summer were offset by the low dissolved oxygen levels and the lower dissolved oxygen saturation value. The lower kinetic removal rates in the winter were compensated by high influent dissolved oxygen concentrations and higher dissolved oxygen saturation values. Since dissolved oxygen penetration was found to be a limiting factor, imposition of high dissolved oxygen concentrations and/or higher dissolved oxygen saturation values will effectively increase the oxygen driving force, increase the active film thickness, and result in increased substrate removal. Thus, although one would expect lower substrate removals during the winter due to suppression of the kientic removal rate, the increased oxygen driving force provides effective compensation. At Edgewater, this resulted in substrate removals similar to those experienced in the summer.

Underflow Clarifier Performance

Beyond the fourth shaft (refer to Figure 3), the RBC/Underflow Clarifier system consists of two distinct sectors, the turnaround sector and the underflow clarifier sector. Tracer analyses indicated that the entire turnaround sector behaved as a

TABLE 4

COMPARISON OF SUMMER AND WINTER PERFORMANCE

		Summer 7/18/77-9/25/77	Winter 12/1/77-2/24/78
Hydraulic Loading Rate $m^3/d/m^2$ (gpd/ft^2)		0.085 (2.08)	0.081 (2.0)
$TBOD_5$ Loading Rate $g/d/m^2$ ($lb/d/1,000\ ft^2$)		11 (2.3)	13 (2.6)
RBC Influent BOD_5 mg/l	T	134	158
	S	97	91
Effluent BOD_5 mg/l	T	28	33
	S	23	24
$TBOD_5$ removal (%)		79	79
$SBOD_5$ removal (%)		76	74
Temperature °C		26.1	11.3
Influent D.O.		1.4	6.1

completely mixed system and that the mixing characteristics in this sector effectively reduced the volume nominally associated with the underflow clarifier from 100 to 75 m^3 (26,000 to 20,000 gallons), or by approximately 25 percent. The nominal surface area, i.e., that which is below the intermediate floor is 72.8 m^2 (784 ft^2). Based on the estimated 25 percent reduction, the available, or effective, surface area becomes 54.6 m^2 (588 ft^2).

Figure 8 presents the correlation of effluent TSS as a function of overflow rate, based on average observed data from each of the major sampling periods. The correlation shown on the Figure, while not particularly uniform, implies an allowable effective clarifier overflow rate between 22 and 26 $m^3/d/m^2$ (550 and 650 gpd/ft^2) to obtain an effluent TSS less than 30 mg/l. At Edgewater this is equivalent to a hydraulic loading rate to the RBC of 0.065 to 0.08 $m^3/d/m^2$ (1.6 to 1.9 gpd/ft^2), assuming an effective surface area of 54.6 m^2 (588 ft^2). The percent removals shown on Figure 7 are based on a fourth stage concentration of 160 mg/l, which was the overall average TSS in the fourth stage during the experimental program.

A solids inventory was maintained throughout the experimental program, and, as expected, the data indicated increasing inventories with increasing BOD_5 removal rates. Figure 9 is a correlation of the total BOD_5 removal rate with total sludge wastage rate (where total sludge wastage is equal to the sum of the pumped waste solids and the effluent solids). The slope of the line is 1.25 lbs sludge/lb $TBOD_5$ removed. Excluding the average influent SS of 125 mg/l, this implies a production rate of 0.3 to 0.4 lb SS per lb of $TBOD_5$ removed.

EVALUATION OF PROCESS DESIGN ALTERNATIVES

Pretreatment

Observations and data gathered in the study indicated a need for pretreatment to remove grit, trash and floatables prior to the RBC system. In the case of Edgewater, high rate primary treatment, with overflow rates between 285 and 370 $m^3/d/m^2$ (7000 to 9000 gpd/ft^2) was found to be adequate, achieving between 20 and 25 percent removal of TSS. Rough screening was necessary to remove large fibrous material which passed through the high rate primary treatment sector.

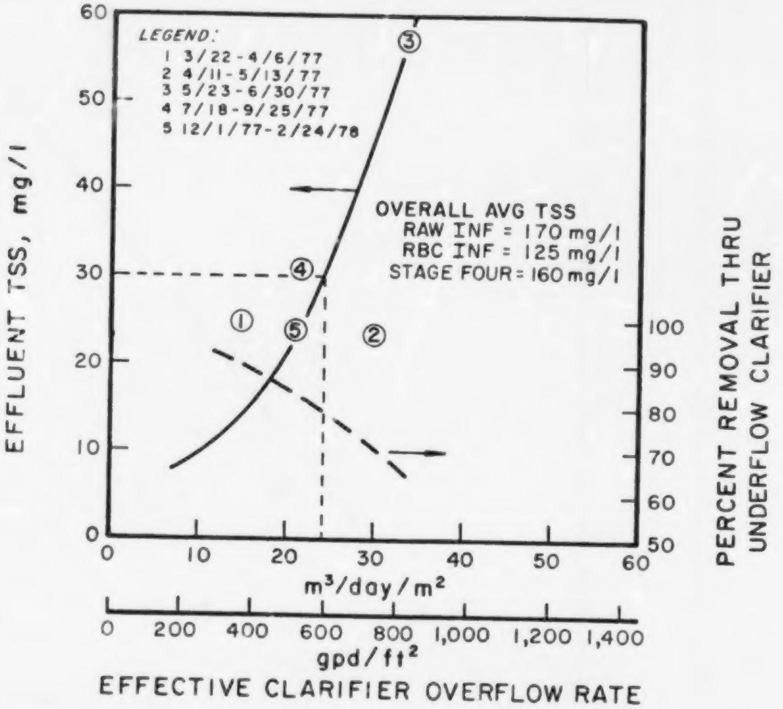


Figure 8. Correlation of effective TSS with effective clarifier overflow rate

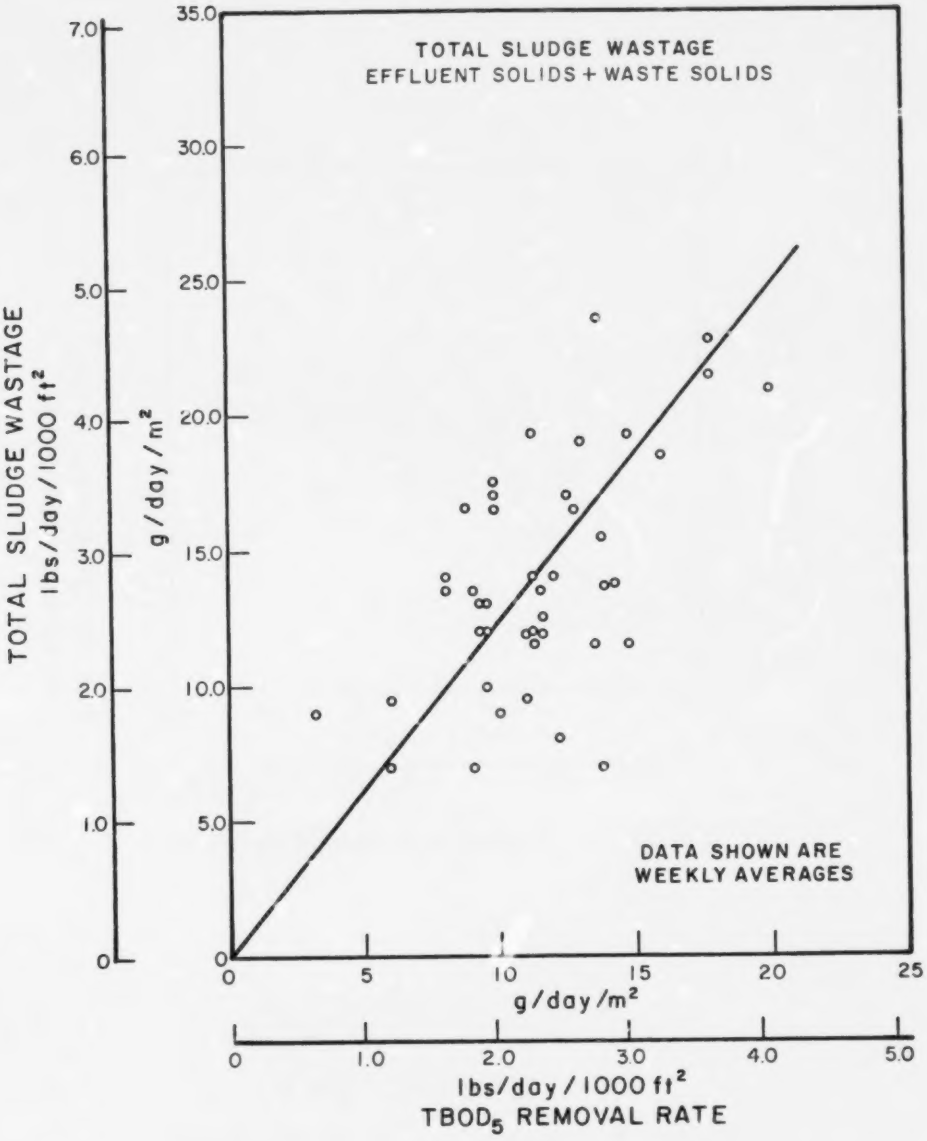


Figure 9. Correlation of total suspended solids wastage and $TBOD_5$ removal

Rotating Biological Reactors

RBC fixed film systems function primarily in the removal of soluble organic material, measurable as soluble BOD_5 (and COD). Thus the design of the system is based on soluble organic loading and soluble effluent organic requirements. The fraction of the $TBOD_5$ influent loading associated with solids will be removed from the system by clarification and these removals are related more to the hydraulic loading of the system. The soluble removals, however, are directly related to biofilm kinetics and the ability of the system to transfer sufficient oxygen. The design sequence assumes, based on the above, that the secondary clarification sector will provide adequate solids removal efficiency and reduce TSS levels to within a desired range. The BOD_5 associated with these solids can be computed from measured BOD_5 to TSS correlations and from this the required effluent soluble BOD_5 can be determined.

Single stage design nomographs were developed on the basis of the kinetic model verifications discussed earlier. These design curves were developed from the system evaluation at Edgewater and, as such, should not be directly applied to the design of systems for treatment of different wastewaters. Application to a particular wsatewater would require that the appropriate kinetic parameters be determined and new design nomographs developed. The curves in this instance are based on an evaluation of a municipal wastewater system and may be useful in preliminary design applications and general process sizing for the treatment of similar wastewaters.

The design of an RBC system should maximize BOD removals in each stage by controlling the BOD loading on the media surface. Maximizing removals in each stage minimizes the total media surface area requirements, thereby minimizing the initial capital expenditure requirements.

Figure 10 presents a series of single stage solutions based on a temperature of 20°C, and an influent DO of 0.0 mg/l. The reaction kinetics described and verified in the previous section were used in the development of the curves where the effluent soluble BOD_5 is determined at the appropriate influent soluble BOD_5 and hydraulic loading rate. The predicted effluent $SBOD_5$

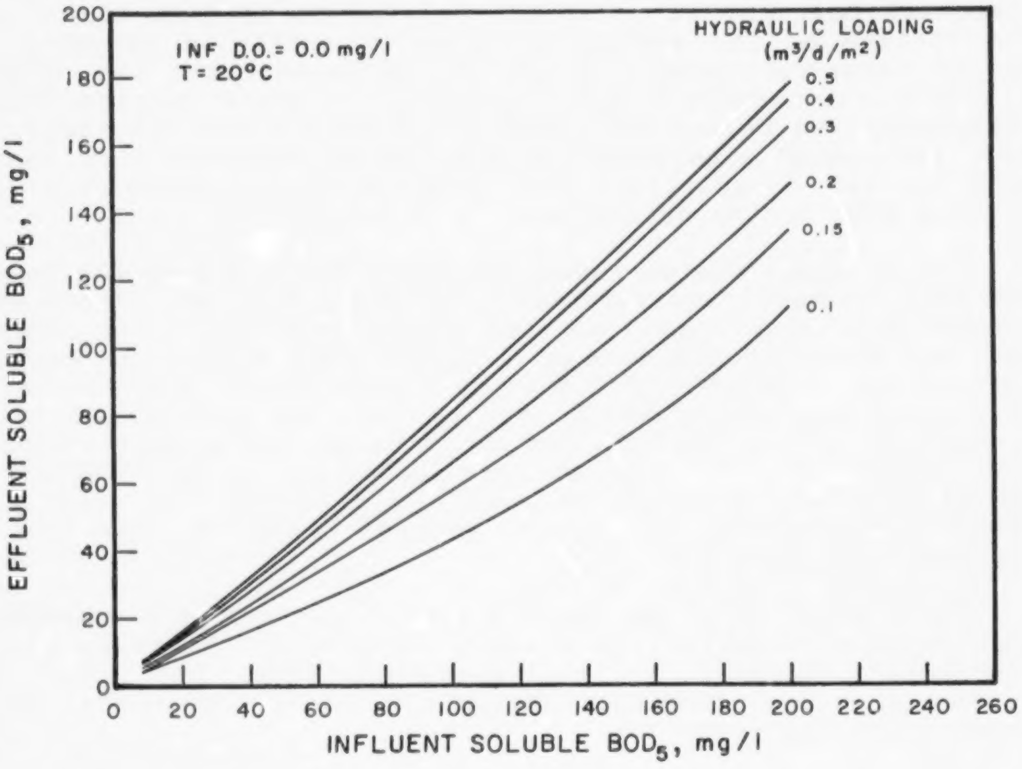


Figure 10. Single stage process design solutions relating effluent $SBOD_5$ to influent $SBOD_5$ and hydraulic loading

concentration from the first stage becomes the influent SBOD$_5$ to the second stage. The iterative use of the design curves allows the prediction of the effluent from a multi-stage RBC system.

Influent Dissolved Oxygen Effects

A second design curve, Figure 11, presents the effect of influent DO on the treatment efficiency of the RBC system. The presence of DO in the influent provides an additional source of oxygen for the biofilm, and may additionally allow a higher concentration gradient, enhancing mass transfer into the biofilm. As indicated on Figure 11, the greater impact occurs at the higher substrate levels. At an influent SBOD$_5$ of 150 mg/l, an influent DO of 6.0 mg/l may allow approximately 12 percent improvement in BOD$_5$ removed in the initial stage.

Secondary Clarification

The pilot study evaluation at Edgewater indicated that a limiting process condition in the operation of the system was the solids removal efficiency accomplished by the secondary clarification zone. The experimental data determined that the maximum effective overflow rate to obtain an effluent SS of 30 mg/l was 26.5 m^3/d/m^2 (650 gpd/ft^2). Thus, the hydraulic loading to the RBC system may be limited by the effective surface area available in the secondary clarification zone. Tracer analyses determined this to be 54.6 m^2 (588 ft^2) in the Edgewater system (a 25 percent reduction from the nominal area of 72.8 m^2). The maximum average flow to the system would therefore be 1,450 m^3/d (0.38 mgd).

The flow of 1,450 m^3/d (0.38 mgd) is equivalent to a hydraulic loading of 0.07 m^3/d/m^2 (1.93 gpd/ft^2) for the Edgewater system. At an influent soluble BOD$_5$ of 90 mg/l, the design curves on Figure 10 project an effluent soluble BOD$_5$ of 19 mg/l. Adding the BOD$_5$ associated with the 30 mg/l TSS, the total BOD$_5$ is projected at 26.5 mg/l. Although this will meet the effluent criteria, the secondary clarifier is effectively limiting the design of the RBC system to 18,270 m^2 (43,000 ft^2) effective media surface area. Denser media, which would allow a higher organic loading, could not be considered since the clarifier would become hydraulically overloaded.

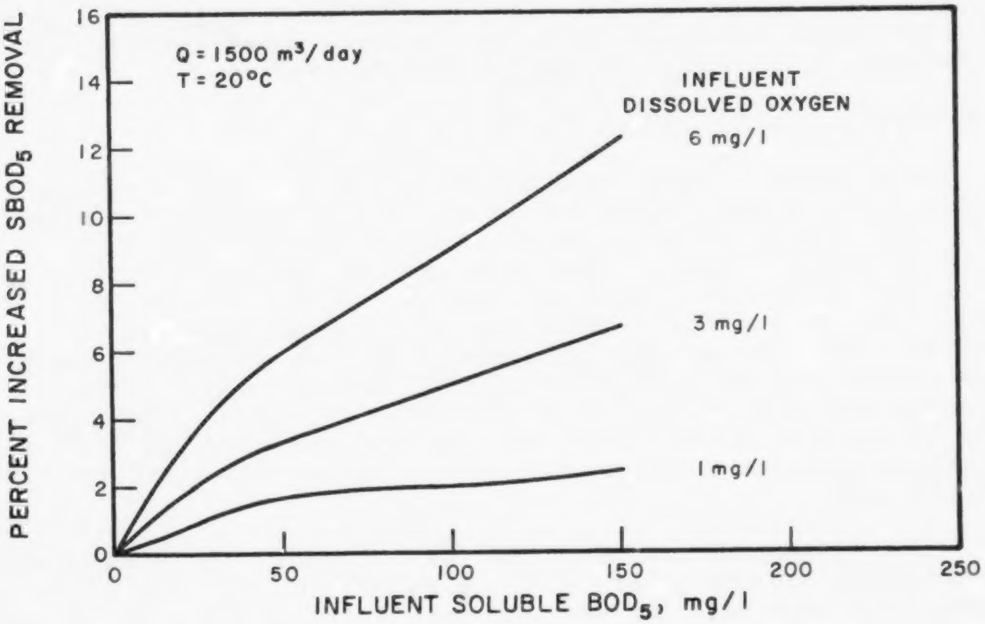


Figure 11. Single stage process design curves relating the effect of dissolved oxygen on $SBOD_5$ removals

To maximize the organic loading to the RBC sector and minimize the RBC surface area requirements, consideration must be given to the design of the underflow clarifier system to accomplish efficient solids capture. This may involve provision of additional secondary clarifiers, the use of chemical addition to improve the efficiency of the existing underflow clarifiers, or the use of rapid sand filters as a final treatment step.

PROCESS DESIGN EXAMPLE FOR THE EDGEWATER SYSTEM

The following example is presented to demonstrate the use of the design curves and to further discuss process considerations relating to the RBC/Underflow Clarifier system. Since the curves are based on the experimental program at Edgewater, the example describes the process requirements to upgrade the existing Edgewater facility to secondary treatment capabilities, based on the present day waste characterization.

The peak monthly average conditions are summarized on Table 5. These were determined to be the controlling conditions in the Edgewater process design.

TABLE 5

PEAK MONTHLY AVERAGES

Flow, m^3/d	13600
(mgd)	(3.6)
Influent	
BOD_5, mg/l (T)	215
(S)	130
TSS, mg/l	260
SBOD Loading, Kg/d	1750
(lbs/d)	(3900)
D.O., mg/l	1-5
Temperature, °C	11-26
Effluent	
BOD_5, mg/l (T)	30
(S)	20
TSS, mg/l	30

Design of the RBC system is based on the loadings to the system, while the clarifier design is considered on the basis of peak flows. The required effluent criteria which apply to Edgewater under the stated influent waste characterization are also presented on Table 5.

Pretreatment

The results of the Edgewater study indicated that the influent to the RBC sector should not exceed 120 to 140 mg/l suspended solids. A single primary clarifier at Edgewater would be used to provide high rate primary treatment for the entire plant flow. Twenty to twenty-five percent TSS removals can be expected. No removal of BOD_5 is assumed through this pretreatment step.

Secondary Clarification

Based on the results of the experimental program, an overflow rate of 24.7 $m^3/d/m^2$ (600 gpd/ft^2) would be necessary to achieve the effluent criteria of 30 mg/l SS. At peak monthly flow of 13,600 m^3/d (3.6 mgd), an effective surface area of 6000 ft^2 is indicated for the Edgewater system.

RBC Organic Removal

Maximum use of the tankage would allow four 4.1 m (13.5 ft) shafts (3.65 m diameter) per tank. High-density media would be installed in all but the first shaft in each tank. Conventional density media would be installed in the first stage. Each shaft would be 0.46 m (30 in) above the water surface.

Figure 12 presents the solutions for a varying number of tanks based on the design curves shown on Figures 10 and 11. For the particular application described on Table 5, the soluble effluent BOD_5 criteria under peak monthly conditions would be met with a total of nine tanks, each with four RBC shafts and a total effective media surface area of 22,600 m^2 (243,000 ft^2) per tank. The total nominal media surface area per tank would be 27,300 m^2 (294,300 ft^2).

Assuming extension of the intermediate floor to provide an effective surface area of 64.6 m^2 (695 ft^2) per tank, nine tanks would provide a total surface area (effective) of 580 m^2 (6,250

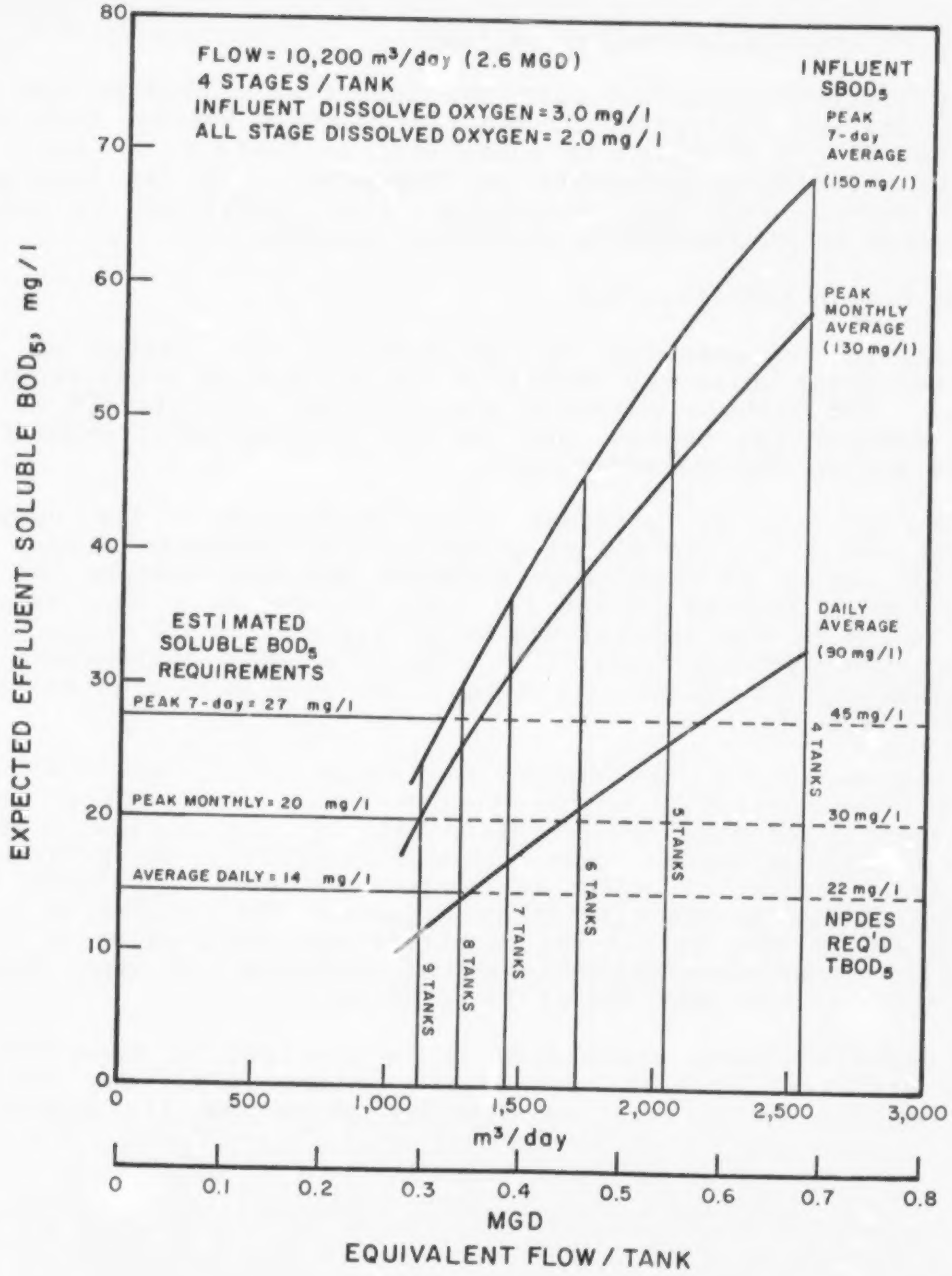


Figure 12. Process design at Edgewater

ft^2). This would be in line with the required secondary clarifier surface area under peak monthly flow conditions.

Summary of Process Design Evaluation

The process design of both the underflow clarifier and the RBC sectors was controlled by the peak monthly average condition at Edgewater. The design is summarized on Table 6. A total of ten tanks would be necessary at Edgewater, one for high-rate pretreatment, and the remaining nine modified or newly constructed as RBC/Underflow Clarifier systems.

Process Design Modifications

Specific improvements can be made in the design of the Edgewater plant which may result in a reduction of total required tankage. Two methods suggested are aeration to DO levels of 5.0 mg/l thoughout the system, and the use of chemical treatment to improve solids capture efficiency.

Aeration can be provided by a supplemental air supply. Although not evaluated directly during the Edgewater study, the potential impact of interstage aeration was simulated by the use of the design Figures 10 and 11. The results were superimposed over results of the initial Edgewater design example (Figure 12) and are displayed on Figure 13. The simulation indicated that provision of interstage aeration alone did not significantly improve the design.

Tests were conducted during the study which indicated that the addition of $FeCl_3$ to the fourth stage mixed liquor would significantly enhance the settleability of the solids. Bench scale tests indicated that it was necessary to provide a sufficient period of agitated contact between the waste and coagulant prior to the clarification zone. The results of these studies showed that within the clarifier operating range of 20 to 25 $m^3/d/m^2$ (overflow rate), chemical addition (20 mg/l $FeCl_3$) would allow an effluent TSS of 15 to 20 mg/l.

A rapid mix zone would need to be provided at Edgewater to assure efficient chemical treatment. Alternatives may involve injecting the $FeCl_3$ solution directly above the air header if

TABLE 6

PROCESS DESIGN SUMMARY AT EDGEWATER UNDER PRESENT CONDITIONS

Process Conditions:

Peak monthly average	
$TBOD_5$ loading	2,120 kg/day 94,670 lbs/day)
$SBOD_5$ loading	1,320 kg/day (2,900 lbs/day)
$SBOD_5$	130 mg/1
Temperature	20°C
Flow (based on loading)	10,200 m^3/day (2.7 mgd)
Influent DO	3.0 mg/1
Flow to clarifier	13,600 m^3/day (3.6 mgd)

Process Design Parameters:

$TBOD_5$ loading rate	10.4 g/d/m^2 (2.1 lb/d/1,000 ft^2)
$SBOD_5$ loading rate	6.5 g/d/m^2 (effective) (1.3 lb/d/1,000 ft^2)
Equivalent hydraulic loading rate	0.05 m^3/d/m^2 (1.2 gpd/ft^2)
Clarifier overflow rate at peak Monthly Hydraulic Flow	23.5 m^3/d/m^2

Process Design (using existing Edgewater tank design):

Total nominal RBC media Surface Area	246,000 m^2 (2.65 x 10^6 ft^2)
Total effective RBC media Surface Area	203,400 m^2 (2.2 x 10^6 ft^2)
Shafts/Tank	4
Total RBC tanks	9
Total Intermediate Floor Surface Area	580 m^2 (6,200 ft^2)

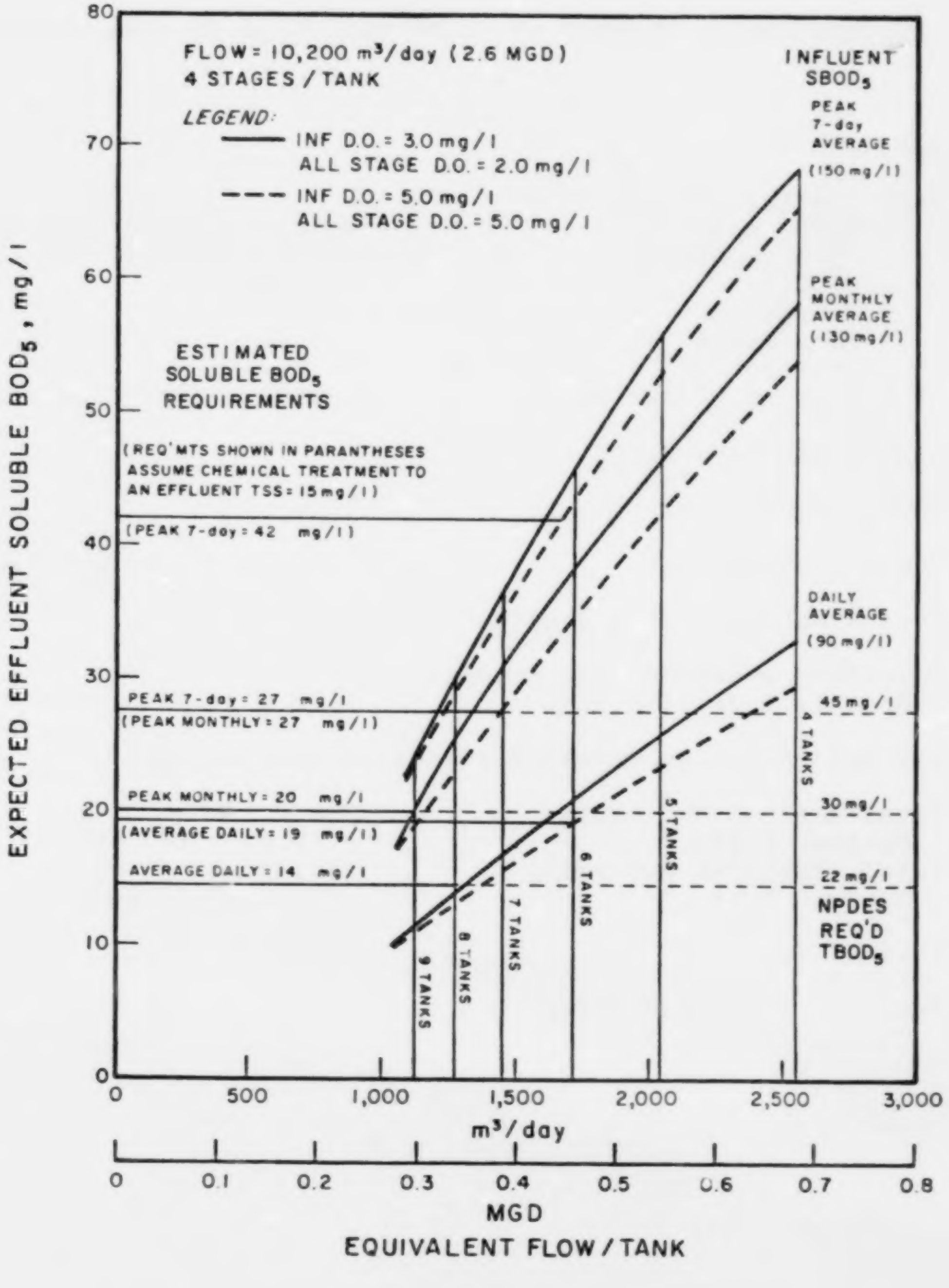


Figure 13. Process design at Edgewater with aeration and chemical treatment

supplemental air is being provided or by installing a separate baffled stage on the extended intermediate floor, with adequate mechanical mixing.

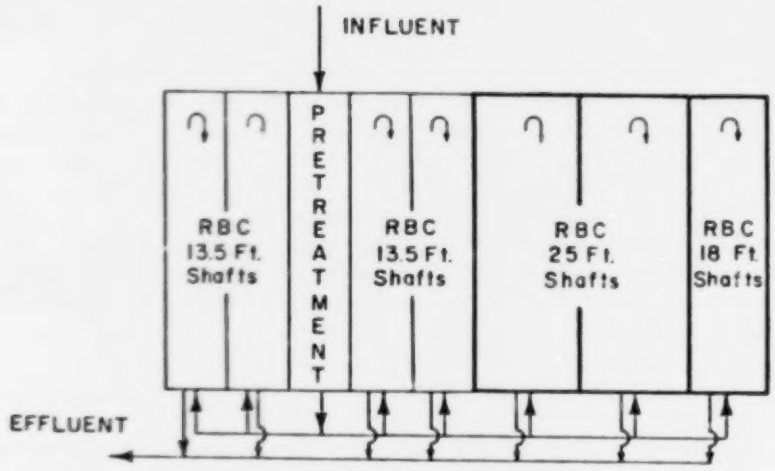
If the effluent solids are maintained at 15 mg/l, soluble BOD_5 effluent requirements change significantly. These are shown on Figure 13. The BOD_5 associated with the 15 mg/l TSS is assumed to be 3 mg/l. Thus the daily, peak monthly, and peak 7-day average $SBOD_5$ requirements become 19, 27 and 42 mg/l, respectively. As shown on Figure 13, the peak monthly condition again governs, but the tankage requirement for the RBC system is now reduced to seven tanks (vs. nine in the initial design), assuming provision of supplemental air.

COSTS

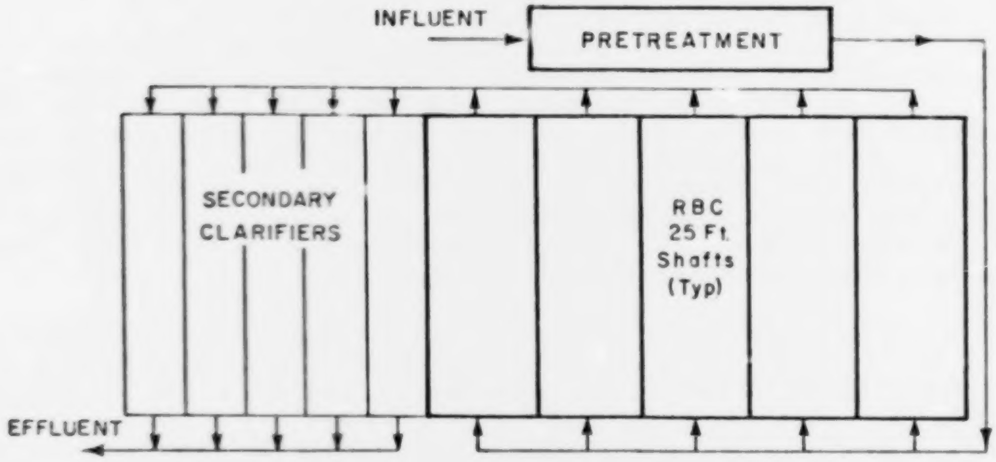
Cost analyses were conducted of alternative design sequences at Edgewater. These costs are based on conditions at Edgewater, including 1977 loading estimates and removal rate coefficients determined during the field program. It is important to realize that costs will be sensitive to these parameters. Thus higher removal rate coefficients would induce lower capital and operating costs.

In the first alternative, one of the existing five primary sedimentation tanks would be converted to a high rate pretreatment tank, while the remaining four would be converted to the RBC/Underflow Clarifier process. new tankage (approximately equivalent to the existing tankage) was then added to provide the requirement for additional surface area in both media and underflow clarification to meet secondary effluent objectives. This alternative is shown in the upper display on Figure 14. The unit cost for this upgrading procedure is estimated to be 0.077 $\$/m^3$ (\$0.29/1,000 gal), considering both operating and maintenance, and amortized captal costs.

A second alternative considered was high rate pretreatment, standard RBC tankage (no underflow clarifier), and utilization of the existing primary tanks for secondary clarification as shown in Figure 14. The unit cost of this scheme is estimated to be $\$0.061/m^3$ (\$0.23/1,000 gal), which is somewhat less than the above RBC/Underflow Clarifier design alternative. Land requirements (included in these costs), however, would be 50 percent higher.


INFLUENT
RBC
13.5 Ft.
Shafts
PRETREATMENT
RBC
13.5 Ft.
Shafts
RBC
25 Ft.
Shafts
RBC
18 Ft.
Shafts
EFFLUENT

CASE 1 - RBC / UNDERFLOW CLARIFIERS
ESTIMATED ANNUAL COST $ 0.29 PER 1,000 GALLONS


INFLUENT
PRETREATMENT
SECONDARY
CLARIFIERS
RBC
25 Ft.
Shafts
(Typ)
EFFLUENT

CASE 2 - SEPARATED RBC UNITS
ESTIMATED ANNUAL COST $ 0.23 PER 1,000 GALLONS

Figure 14. Mechanical drives schematic layout.

SUMMARY

1. The RBC/Underflow Clarifier concept was demonstrated to be an effective secondary treatment process, capable of meeting NPDES secondary treatment effluent requirements of 30 mg/l $TBOD_5$ and TSS, or 85 percent $TBOD_5$ and TSS removal, whichever provides the greater degree of treatment.

   The peak monthly loading at Edgewater controlled the process design. The influent peak monthly total BOD_5 was 215 mg/l, with a corresponding soluble BOD_5 equal to 130 mg/l. Limiting organic loadings for the RBC sector were determined to be 10.4 g $TBOD_5/d/m^2$ (2.1 lb/d/1000 ft^2) and 6.5 g $SBOD_5/d/m^2$ (1.3 lb/d/1000 ft^2).

2. Pretreatment of the raw wastes was required throughout the study period to remove grit, trash, and floatables. Rough screening was necessary to remove large fibrous material which passed through the high rate primary treatment sector.

3. Tracer analyses indicated that each RBC stage, as defined by baffle placement, behaved as a completely mixed tank. The mixing characteristics of the turnaround sector reduced the effective volume of the clarifier by approximately 25 percent.

4. The overall organic removal efficiency of the RBC system was limited by oxygen availability, as determined by use of an RBC mathematical kinetic model. Oxygen utilization curves developed from COD balances and the model indicated that the system reached a limiting condition in its ability to transfer oxygen at the higher influent organic loading rates.

5. The overall seasonal effects were minimal based on the evaluation of the system under summer and winter conditions. The temperature differential experienced was 15°C. Although temperature affects several mechanisms involved in the kinetics of the fixed film process, the minimal overall impact experienced over this large temperature differential was due to compensating influences of the various parameters affected by temperature.

6. Pre-aeration was investigated using the kinetic model. Since the system at Edgewater is characterized by decreasing organic load with progressive staging, the provision of pre-aeration to the influent of the RBC/Underflow process would not have a significant impact on removal efficiency. Interstage aeration would achieve greater substrate removals.

7. Underflow baffles effectively stage the RBC system into a series of completely mixed tanks. Baffling also created higher velocities along the intermediate floor and minimized solids accumulation. At an initial baffle clearance of 15.2 cm (6 in), velocities were not sufficient to prevent considerable accumulations on the floor. Reduction of the baffle clearance to 5 cm (2 in) effectively prevented further solids accumulation.

8. A linear correlation of total suspended solids wastage as a function of $TBOD_5$ loading to the system was determined. On average, biological solids growth was estimated to be 0.38 g SS_p/g BOD_5 removed.

9. A correlation of effluent TSS and underflow clarifier overflow rate was constructed on data collected over the entire program. The correlation implies an allowable overflow rate between 22 and 26 $m^3/d/m^2$ (550 and 650 gpd/ft^2) to obtain an effluent TSS less than 30 mg/l.

10. Chemical addition studies showed that ferric chloride addition to the fourth stage effluent would effectively improve solids settleability.

11. Cost analyses were conducted of alternative design sequences at Edgewater. The unit cost for the RBC/Underflow Clarifier upgrading procedure was estimated to be \$0.08/m^3 (\$0.3/1000 gal).

ACKNOWLEDGEMENTS

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UPGRADING TRICKLING FILTER EFFLUENTS
WITH A RBC SYSTEM

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INTRODUCTION

Many secondary treatment plants in the United States were built two decades ago. The treated effluents of these plants cannot meet the current effluent quality standards. Several alternatives can be taken to improve the effluent quality. Among these alternatives are: (1) construction of a new treatment plant for the existing and future loads; (2) expansion of the existing plant; and (3) addition of an upgrading retrofit system to the existing plant. The last alternative is deemed desirable because of lower cost requirements and added operational flexibilities.

This paper describes a pilot plant study using a rotating biological contactor (RBC) system to remove organic as well as to induce nitrification from a trickling filter effluent. The application of a RBC treatment system to upgrade secondary effluent is relatively new in the United States. Only a few plants have been in operation for more than one year. Lagnese (1978) described the use of a RBC system operated in parallel as well as in series with trickling filters. According to his findings, it was suggested that soluble BOD (SBOD) instead of total BOD should be used in the RBC process design. Reh (1977) also suggested the same approach since they considered that particulate BOD was removed by various physical phenomena and its removal was not really a function of the biological oxidation activity of the biomass. Lagnese (1978) also reported in his study a large portion of the effluent BOD from the RBC process following a trickling filter was nitrogenous BOD (NBOD). Sullivan (1978) reported the operation of a few treatment plants using RBC systems to upgrade secondary effluent qualities. Since only a scanty description of the operation and a meager amount of data were presented in his report, the RBC performance for such application cannot be properly evaluated.

PILOT PLANT INSTALLATION AND OPERATION

The pilot plant module was a standard 0.5-m diameter, 4-stage RBC unit manufactured by Autotrol Corporation, Wisconsin. In operation, approximately 40 percent of the 23.3 m^2 total contact surface area was submerged in wastewater. The single speed motor furnished for the module unit was replaced with a variable speed motor so that the effects of rotating velocity on unit performance could be studied. Chemical supplements could be added, if needed, from reservoirs holding the prepared chemical solutions to the pilot unit using chemical pumps. A clarifier (polyethelene tank) of 0.51-m diameter was equipped with a 0.5-m long V-notched overflow weir. The effluent from the clarifier was discharged into the chlorination chamber of the existing trickling filtering plant. The entire pilot unit, with the exception of the clarifier, was housed in a sheet-metal shed.

It took 9 days to develop a thin layer of biomass covering the entire outside of the media. Both the wastewater temperature and the ambient temperature during this startup period were at 11°C. Commencing on January 10, 1978, the pilot plant was run continuously in a 7-month span. Random grab samples were taken from the pilot plant influent, effluent, and sometimes from each of the four stages for chemical and biological analyses. In addition the amount of biological slime on the rotating media was determined from time to time by scraping the culture from a 20-cm^2 surface area. Sludge volume index was determined from 30-min. settling in an Imhoff cone. Graduate cylinder batch settling test was used to determine the sludge settling velocity as well as the solid flux from which the proper size of a clarifier-thickener could be determined.

RESULTS AND DISCUSSION

Soluble BOD_5 Removal The effluent $SBOD_5$ concentration obtained from the pilot plant operation is illustrated in Figure 1. The soluble nitrogenous BOD_5 ($SNBOD_5$) line is obtained by substracting the soluble carbonaceous BOD_5 ($SCBOD_5$) from the $SBOD_5$. When the effluent contains 15 mg/l $SBOD_5$ and an average of 15 mg/l suspended solids (the later contributes from 12 to 18 mg/l of BOD_5 to the effluent, averaging 15 mg/l), the equivalent total BOD_5 of the effluent is 30 mg/l which meets the current standards of secondary effluent. For this effluent of 15 mg/l $SBOD_5$, Figure 1 shows that an influent $SBOD_5$ loading up to 7.8 g/m^2-day can be applied. Assuming a value of 10 mg/l of $SBOD_5$ allowed in tertiary treatment effluents, the influent $SBOD_5$ loading would be reduced to 3.5 g/m^2-day.

To compare the RBC performance in this study with that of a similar study by Lagnese (1978), Table 9 in his report shows that on the average an effluent $SBOD_5$ of 16 mg/l (range 11 to 21 mg/l) was obtained for a hydraulic loading of 0.04 m^3/m^2-day and an influent $SBOD_5$ of 60 mg/l. The influent $SBOD_5$ loading was therefore 7.57 g/m^2-day. Figure 1 of this study shows an average of 15 mg/l effluent $SBOD_5$ at the same loading and therefore the results of these two studies are comparable.

The removal of $SBOD_5$ versus influent $SBOD_5$ loading is shown in Figure 2. It can be seen that the removal in this study (Curve A) was between 55 to 60 percent at an influent $SBOD_5$ loading of 4.88 g/m^2-day or above. Below this loading the removal percentage was lower. Data were taken from Lagnese's study and plotted as Curve B for comparison. The removal in Lagnese's study was higher by a few percentage points possibly because the influent $SBOD_5$ concentration at 60 mg/l was higher than the average of 25 mg/l in this study. Also shown are Curve C (data taken from Reh (1977)) and Autotrol Corp Design Manual (1972) for RBC treatment of primary effluent with higher $SBOD_5$ concentrations) and Curve D (Autotrol prediction model curve, with low $SBOD_5$ concentration at 25 mg/l). The evidence thus far indicates that the $SBOD_5$ removal percentage obtained by RBC treatment decreases with decreasing influent $SBOD_5$ concentrations.

Two reasons can be offered to explain the lower percentage removal of $SBOD_5$. Trickling filters remove the more readily degradable fraction of the wastewater, leaving the more resistent fraction to the RBC. Also the low concentration of the influent $SBOD_5$ could result in a depletion of BOD within the biofilm, a phenomenon known as hindered mass transfer. Famularo (1978) have found that low BOD concentrations in the liquid film coating the biofilm result in a depletion of BOD within the biofilm, leading to inactive biofilm and inefficient BOD removal.

Soluble Nitrogenous BOD_5 Figure 1 shows that the SCBOD_5 of the RBC effluent was quite steady, varying from 2 to 8 mg/l when the influent SBOD_5 loading varied significantly from 0.49 to 7.8 g/m^2-day. Within the same range of SBOD_5 loadings, however, the effluent also contained from 1 to 13 mg/l SNBOD_5. As a result the effluent SBOD_5 concentration was increased by 50 to 160 percent. Calculations using data presented in Table 5 by Lagnese (1978) show that the effluent SBOD_5 was increased by nitrogenous BOD_5 by 33.3 to 115 percent in the RBC unit. If nitrogenous BOD_5 can be excluded from consideration in the secondary effluent standards, the allowable influent SBOD_5 loading can be increased to 61 g/m^2-day by extrapolation from Figure 1.

Effects of Low Temperature and Hydraulic Shock The RBC pilot plant was tested in a low temperature series with the wastewater temperature ranging from 5 to 10°C. Figure 3 shows data points relating the effluent SBOD_5 concentration to the influent SBOD_5 loadings. Also presented in Figure 3 are data points collected from the hydraulic shock series, using 120 to 220 percent of a moderate hydraulic load of 0.045 m^3/m^2-day and at a low influent SBOD_5 concentration.

The two regression lines overlap and indicate that the effluent SBOD_5 increases very rapidly with increasing influent SBOD_5 loadings under the stress of low temperature or hydraulic shock. By keeping the same 15 mg/l of allowable effluent SBOD_5 in Figure 1 and 3 and compare the allowable influent SBOD_5 loadings, a correction factor of 0.41 is obtained (3.2 versus 7.8 g/m^2-day). It is worth noting that this correction factor 0.41 was for the low temperature range 5 to 10°C and for the RBC unit receiving low SBOD_5 concentrations and consequently maintaining a low biological population.

Temperature coefficients $\theta = 1.03$, 1.04, and 1.06 are often used respectively for activated sludge, trickling filter, and lagoons. It is noted that the revised design manual by Autotrol (1978) suggests a temperature correction factor of 0.76 for a low temperature of 7.2°C which was the average of the low temperature series in this study. The temperature coefficient $\theta = 1.05$ applies to the condition from which Autotrol developed their correction factor. A RBC unit, being a very efficient heat transfer system according to Murphy (1975), and treating a low BOD wastewater with a low biological population in the present study, would be affected more by temperature similar to lagoons. Because of the unique application of RBC in this study, it is suggested that a temperature coefficient $\theta = 1.07$ be used, or a correction factor of 0.42 which fits very closely the data presented in this study.

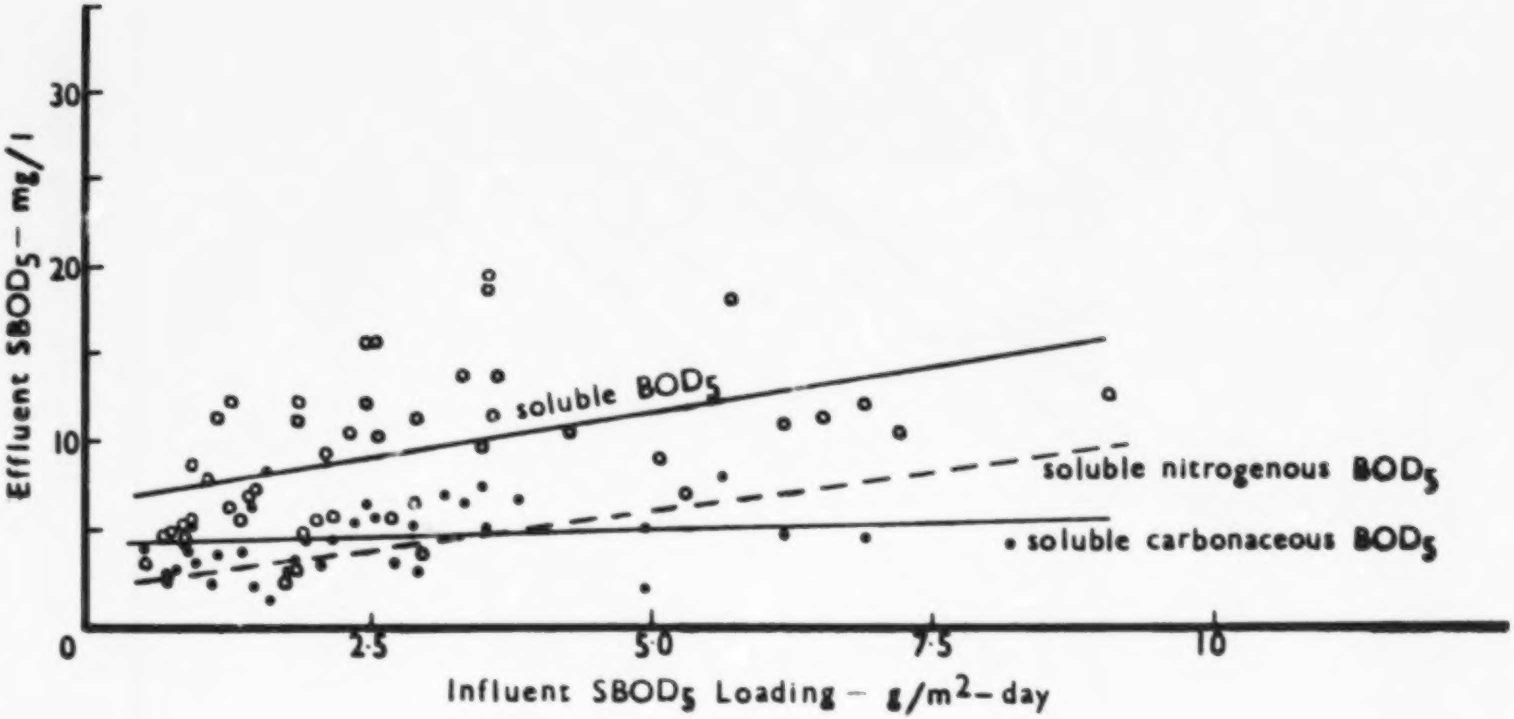

FIGURE 1. Relationship between effluent soluble BOD_5 and influent soluble BOD_5 loading.

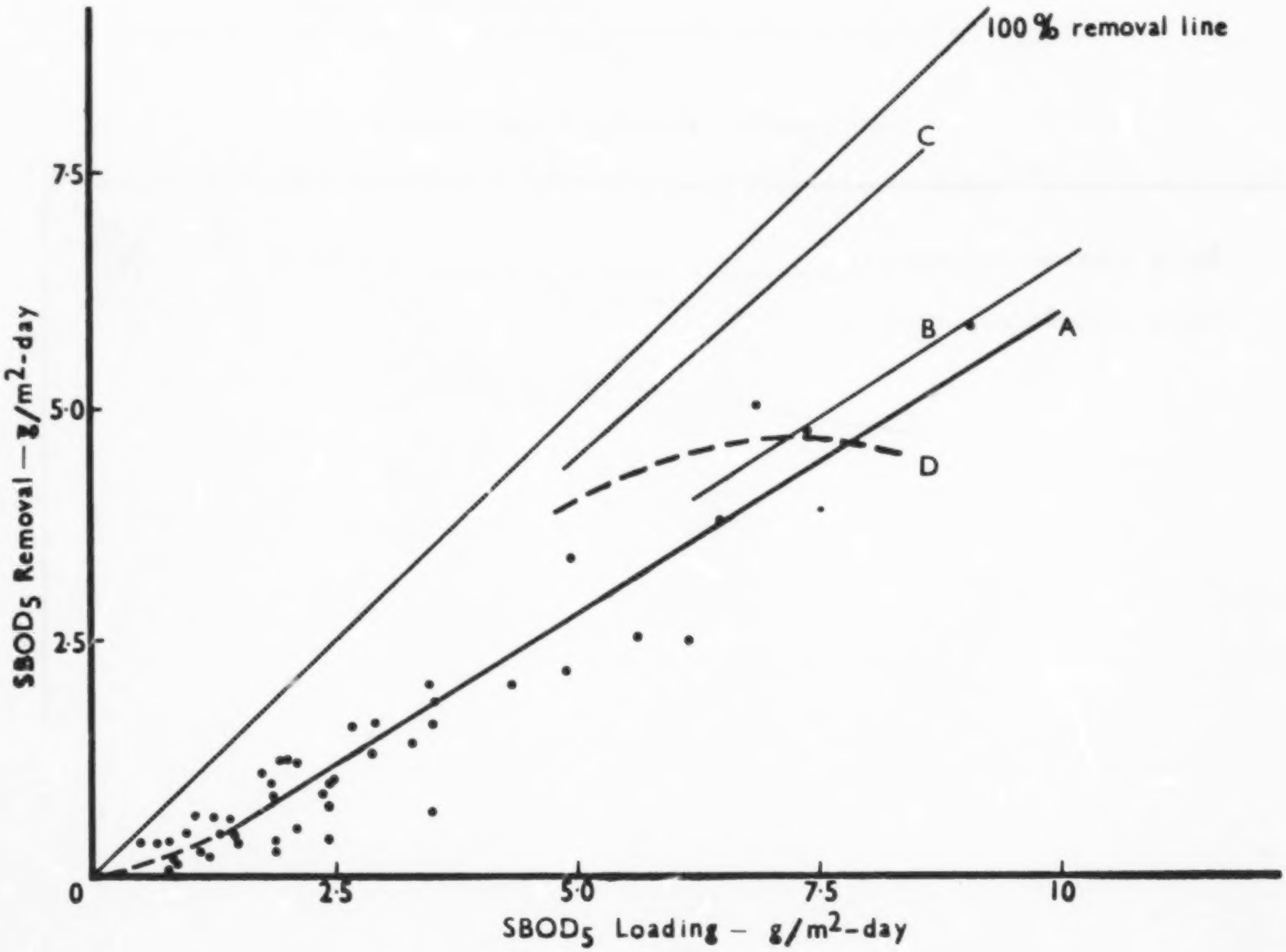

FIGURE 2. Relationship between soluble BOD$_5$ removal and loading.

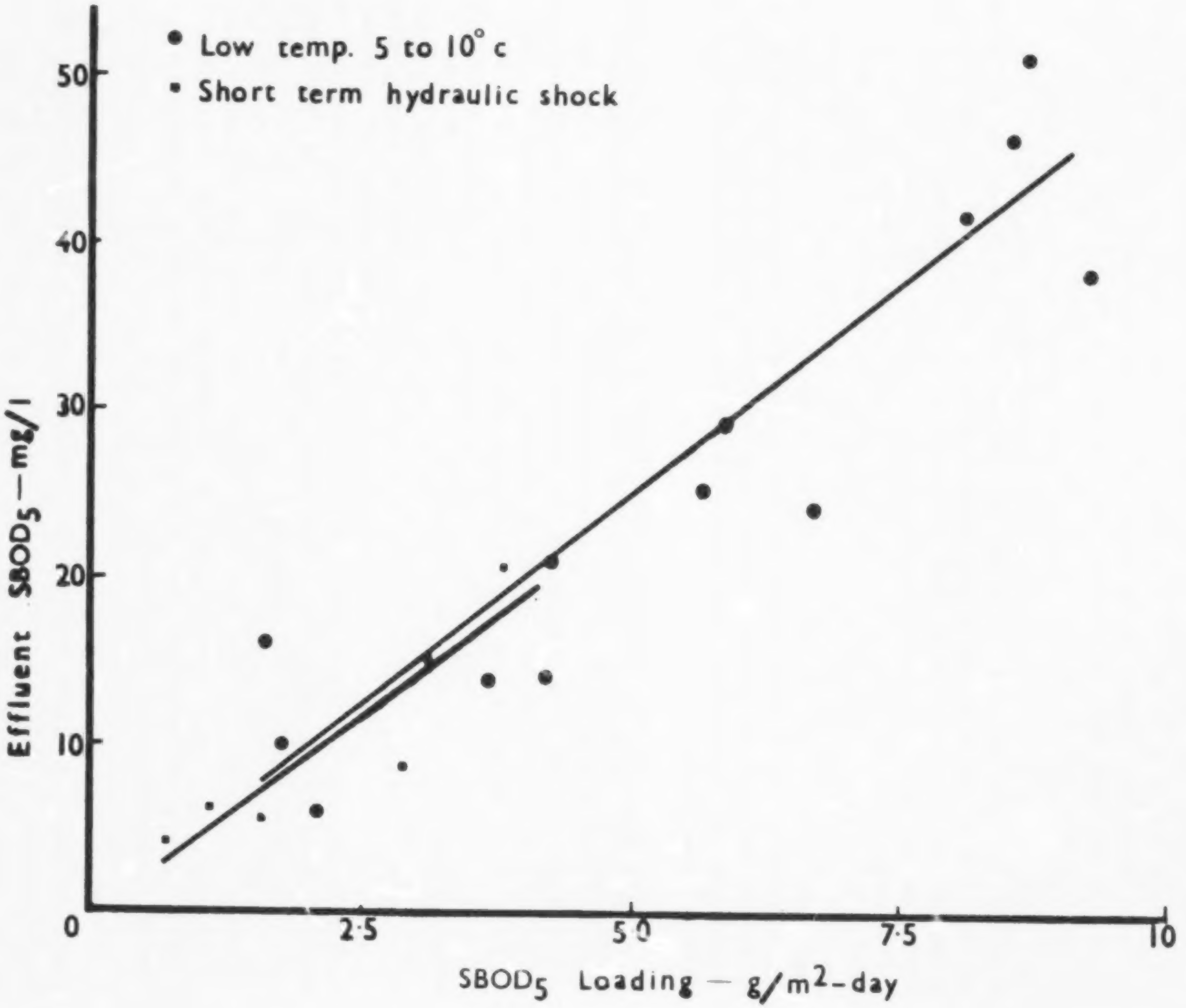

FIGURE 3. Effluent soluble BOD_5 at low temperature or short term hydraulic shock loading.

With hydraulic shock loadings in effect, the shortened detention time (44 to 25 minutes) and the low F/M ratio magnified the hindered mass transfer effect. It was observed that the thickness of the biofilm on the RBC media was reduced slightly. In other words, the RBC unit began to drift to match the population to the low soluble BOD_5 concentration.

Effects of Organic Shock To simulate organic shocks, loadings from 4.3 to 15.4 g $SBOD_5/m^2$-day (124 to 444 percent of the average load) were applied. The highest $SBOD_5$ concentration of the RBC influent was 72 mg /l which was considerably higher than the average of 25 mg /l. Figure 4 shows that if there was a discernable effect, the shock organic loading actually improved the treatment performance slightly. The result is not at all surprising since RBC is known to be able to handle high BOD loadings successfully with good removal. From 17 to 28.3 g total BOD_5/m^2-day can be removed without any adverse effect according to Lagnese (1978) and Poon (1979). Data reported by Reh (1977) show no adverse effect at 17 g $SBOD_5/m^2$-day. Apparently the biofilm can keep up with the higher loadings by removing the $SBOD_5$ at the rate it enters the RBC unit. It is worth noting that nitrification activity is stronger at lower organic loadings and consequently the masking effect of nitrogenous BOD is more pronounced. This explains why a better performance might be obtained at higher $SBOD_5$ loadings.

Soluble NH_3-N Removal and Effects of Low Temperature and High Organic Loading Data shown in Figure 5 indicates that the effluent soluble NH_3-N concentration increases with increasing influent soluble NH_3-N concentration. A similar relationship was found by Antonie (1971). Other studies by Reh (1977) and Reid (1974) also imply similar results. The rate of nitrification decreases with temperature decrease throughout the range from 30 to 5°C, according to EPA (1973). Murphy (1975) reported a good fit of nitrification rate data in their RBC study, using the Arrhenius relationship over a temperature range of 7 to 25°C. The present study was started in winter and nitrification was not very well established throughout the winter and early spring. The low temperature of the wastewater ranged from 5 to 10°C (average 7.7°C) as opposed to the average of 18.8°C for the normal condition in Figure 5. By comparing curves A and B, it can be seen that the effluent soluble NH_3-N concentration was approximately 38 percent higher than that under normal condition for any influent loading within the range indicated in Figure 5.

High organic shock was simulated in this study by applying concurrently a high hydraulic loading and a moderate BOD_5 concentration. High hydraulic loadings from 0.18 to 0.245 m^3/m^2-day resulted in short detention times from 43 to 32 minutes. According to EPA (1973) 60 minutes would be required

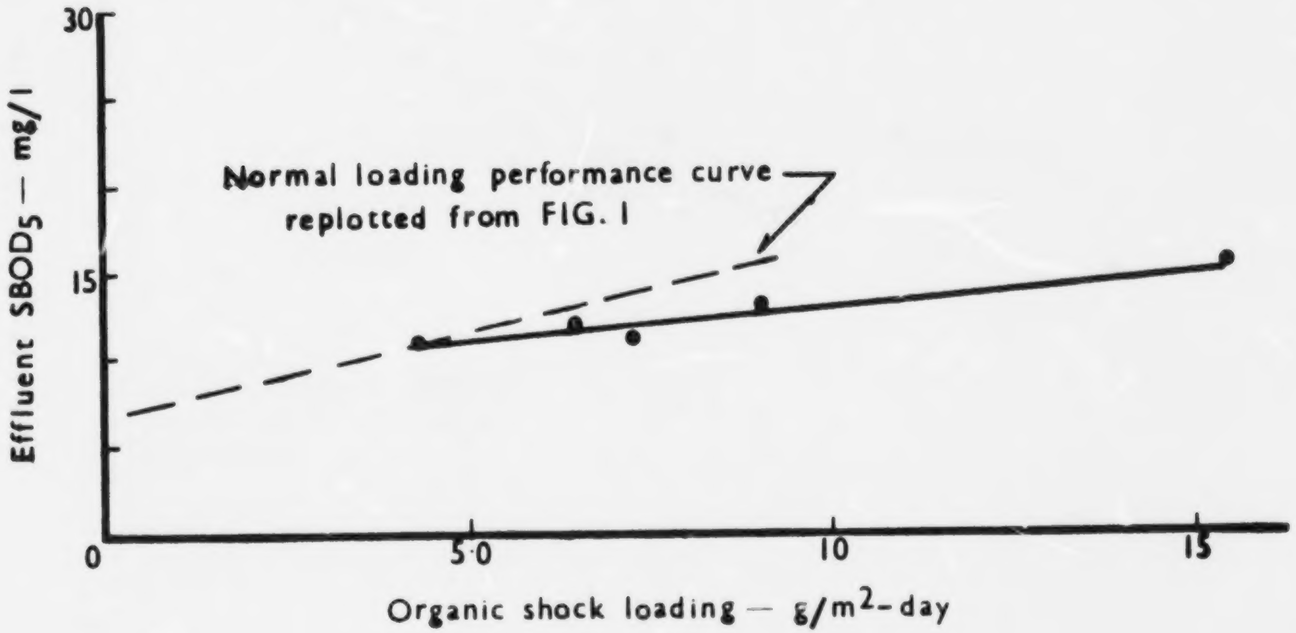

FIGURE 4. Relationship between effluent soluble BOD_5 and influent organic shock loading.

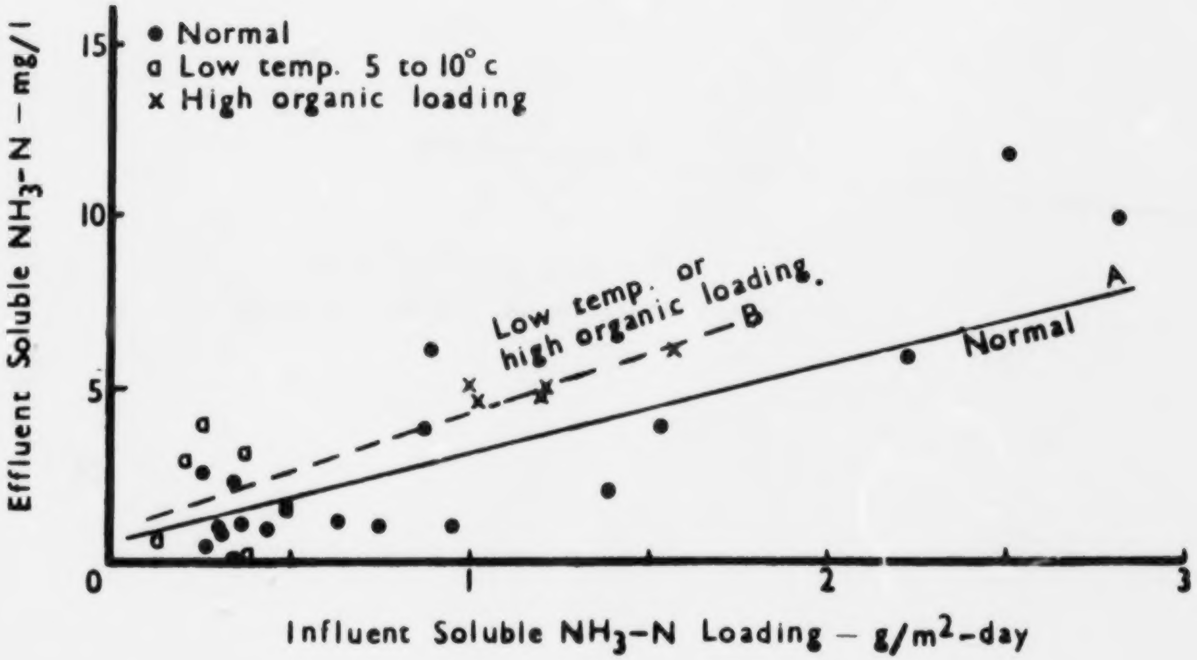

FIGURE 5 . Relationship between effluent soluble NH_3-N concentration and influent soluble NH_3-N loading.

for nitrification, removing 25 to 30 mg/l NH_3-N to 2.5 to 5.0 mg/l in a suspended growth system. For complete NH_3-N removal, 1.5 hours are required at pH 8.5 and 7 hours at pH 6.5. Using a RBC as a secondary treatment unit, Antonie (1971) reported that approximately 60 minutes would be required to accomplish 85 percent NH_3-N removal. At higher loadings a RBC unit would need a minimum of 1.5-day detention time for nearly complete NH_3-N removal, according to Lue-Hing (1974). Another unfavorable factor in this study was the higher BOD concentration in the RBC influent. Bacteria that remove carbonacous BOD grow at a much faster rate than nitrifiers. Consequently they outgrow and reduce the nitrifiers in the RBC unit since no biological sludge is recycled. The BOD limit in RBC systems for successful nitrification is 30 mg /l according to Antonie (1972), and 14 to 20 mg/l according to Antonie (1971). In this study series, the influent total BOD_5 ranged from 35 to 131 mg /l, averaging 89 mg /l. The short detention time coupled with the higher BOD concentration resulted in the reduction of NH_3-N removal as shown in Curve B, Figure 5.

Nitrification Kinetics The soluble NH_3-N removal increased with the influent soluble NH_3-N loading as illustrated in Figure 6. The removal was approximately 50 percent under normal condition (0 to 0.73 g/m^2-day loading), and 83 percent at higher loadings. At low temperatures or high organic shocks (concurrent high hydraulic loading and moderate BOD concentration), the removal was reduced to 20 percent.

The removal term in Figure 6 expressed in g soluble NH_3-N/m^2-day can be called the unit rate of nitrification, similar to that defined by Murphy (1975) as mg TKN/m^2-hr. removal. The temperature effect on the unit nitrification rate was (1-20%/50%) = 60% reduction when the temperature was dropped from 18.8°C (average of normal condition) to 7.7°C (average of low temperature series). Estimated from Figure 4 of Murphy's work (1975), The temperature effect on the unit nitrification rate was 55 percent reduction for the same temperature drop. For a suspended growth system. a reduction of 63 percent of the unit nitrification rate can be estimated from Figure 1-13 of the EPA publication (1973). It thus appears that the temperature effect on the unit nitrification rate is relatively the same for both RBC and the suspended growth systems.

On several occasions NH_3Cl was added to the RBC influent to simulate a NH_3-N overloading situation. Alkalinity in the form of sodium carbonate was also added in addition to NH_3Cl in one occasion. The performance data in this series were included in preparing Figure 7 covering a wide range of soluble NH_3-N loadings. Figure 7 suggests that S-shaped or logistic-S curves best fit the data, expressed as:

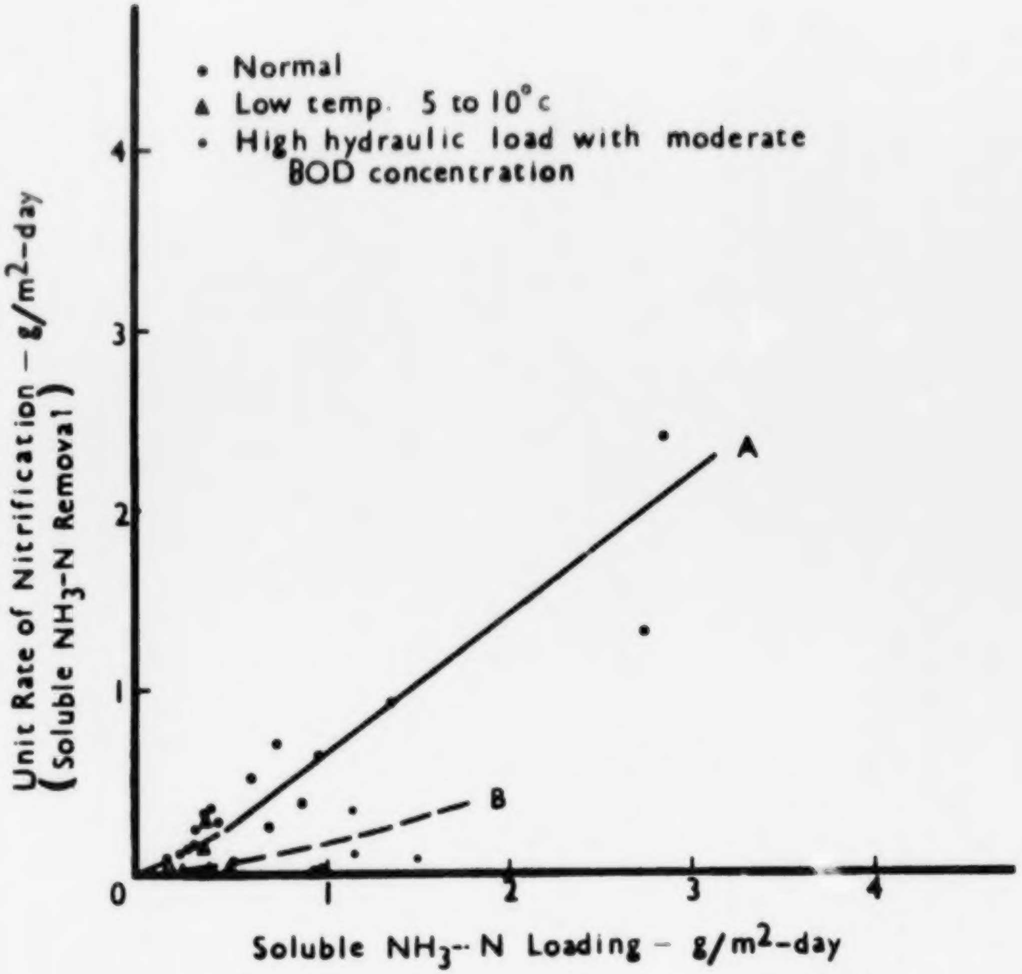

FIGURE 6. Relationship between soluble NH_3-N removal and influent soluble NH_3-N loading.

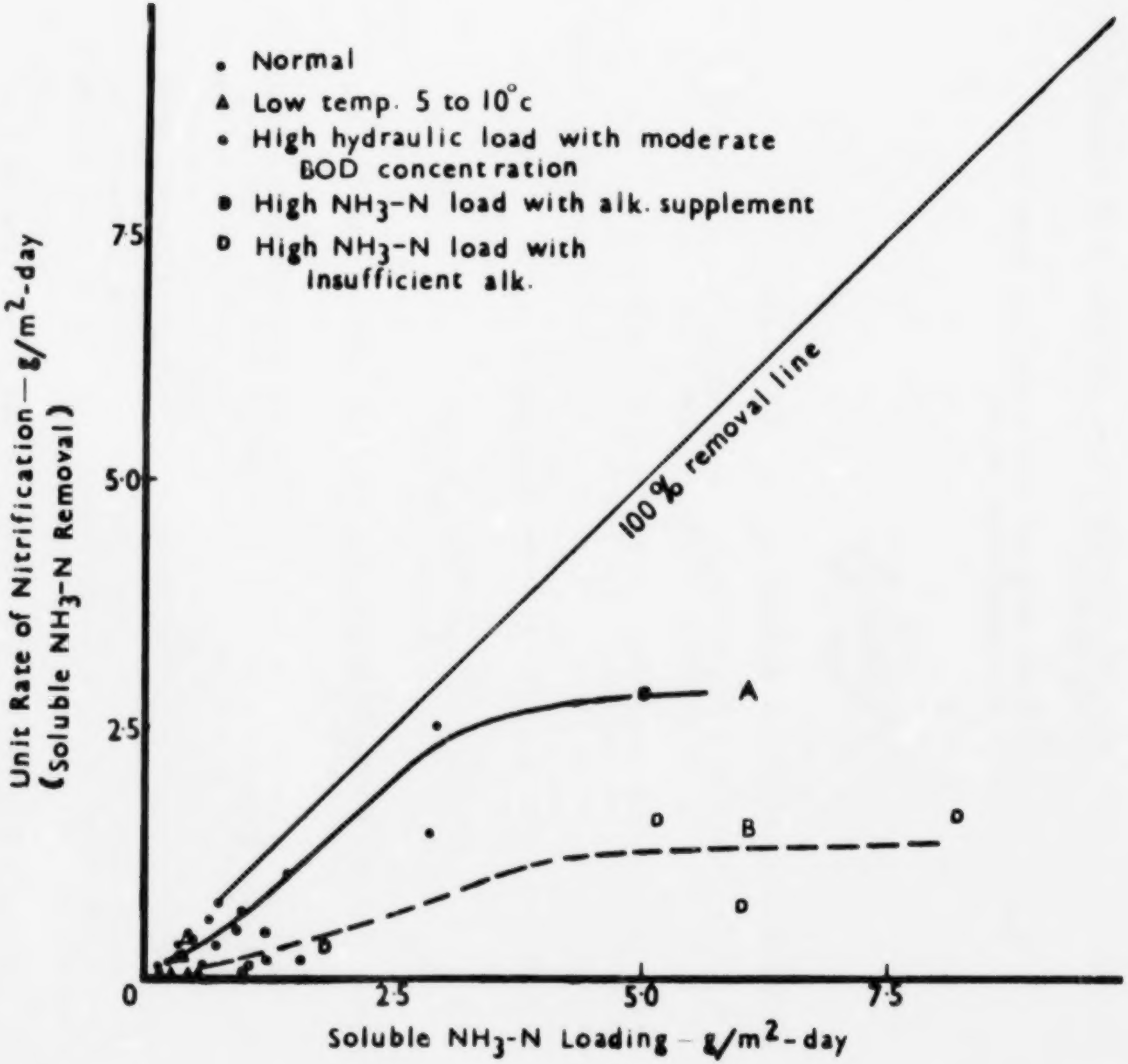

FIGURE 7. Soluble NH_3-N removal at various loadings.

$$R = \frac{R_{max}}{1 + m \cdot e^{b \cdot L}} \tag{1}$$

in which R = unit rate of nitrification, soluble NH_3N removal in lb./1000 ft^2-day or g/m^2-day

R_{max} = maximum unit rate of nitrification, same unit, m and b are constants, where

$$m = \frac{R_{max} - R_o}{R_o}$$

$$b = \frac{1}{n} \ln \frac{R_o}{R_1} \frac{(R_{max} - R_1)}{(R_{max} - R_o)}$$

in which n = intervals between L_o, L_1, L_2 loadings

R_o, R_1, R_2 = soluble NH_3-N removal at loadings L_o, L_1, and L_2, all in lb/1000 ft^2-day or g/m^2-day

Based on the data presented in Figure 7 and for n = 0.3 lb/1000 ft^2-day (loadings 0.1, 0.4 and 0.7 lb/1000 ft^2-day and corresponding removals at 0.05, 0.285, and 0.516 lb/1000 ft^2-day) the developed constants are m = 10.28, and b = 7.84 (Curve A). Consequently the logistic-S curve applicable to this study is:

$$R = \frac{0.564}{1 + 10.28e^{-7.84(\Delta L)}} \tag{2}$$

where ΔL = difference between any applied loading and 0.1 lb/1000 ft^2-day.

A logistic-S equation was found to fit the nitrification data in a natural stream, according to Wezernak (1968). The equation was developed with both nitrifying bacteria and substrate exerting a concentration effect on the rate of nitrification. Such equation would be most applicable to an environment when both nitrifying organisms and NH^3-N concentrations are low. The low soluble NH_3-N loadings and the thin biofilm developed on the RBC media in this study justify the use of this logistic-S equation.

Curve B in Figure 7 was developed for conditions of low temperature, high organic loading, or NH_3-N overloading with insufficient alkalinity in the influent. Curve B can be expressed by the following equation:

$$R = \frac{0.243}{1 + 13.29e^{-6.44(\Delta L)}} \qquad (3)$$

Comparing Curves A and B, there is a reduction of 57 to 75 percent of the unit nitrification rate within the range of influent loadings of 0 to 1.1 lb soluble NH_3-N/1000 ft^2-day. While both kinetic equations 2 and 3 are considered adequate in predicting the unit rate of nitrification, the constant values m and b as well as the R_{max} values were developed specifically for this study. Other RBC plants of similar application should develop specific values of m, b, and R_{max} for their own use.

Biofilm Characteristics and Sludge Synthesis The biofilm thickness was highly uneven at all 4 stages. Thickness of less than 0.5 mm to 2.0 mm was observed at the first stage. The difference of thickness was less significant at the other stages but nevertheless clearly visible. The average growth of the 4 stages were 1.08, 0.79, 0.52 and 0.44 g/1000 cm^2 disc surface area.

The amount of sludge synthesis was determined indirectly by substracting the 4th stage MLVSS concentration from the influent MLVSS concentration and dividing the difference by the $SBOD_5$ removed. The calculated value is 0.28 g MLVSS/g of $SBOD_5$ removed. This value is low compared to the average of 0.5 g/g BOD_5 removed, indicating that RBC as an upgrading retrofit unit has a much lower biomass synthesis. This low synthesis value reflects a possible long sludge retention time in the RBC unit.

Sludge Retention Time Unlike a complete-mix suspended growth system, the sludge retention time of the biofilm can only be determined indirectly by setting up a mass balance equation as follows:

$$\begin{array}{c}\text{change of}\\\text{film biomass}\end{array} = \begin{array}{c}\text{4th}\\\text{stage}\\\text{MLSS}\end{array} - \begin{array}{c}\text{influent}\\\text{MLSS}\end{array} - \begin{array}{c}\text{soluble}\\\text{BOD}\\\text{removed}\end{array} \times \begin{array}{c}\text{growth}\\\text{yield}\\\text{coefficient}\end{array}$$

If the value of the change of film biomass calculated is positive, solid sloughing takes place. A negative change means increase of biofilm thickness and therefore no sloughing. The growth yield coefficient was calculated previously to be 0.28 g MLVSS/g $SBOD_5$, or 0.31 g MLSS/g $SBOD_5$ using a ratio of MLVSS/MLSS = 0.9 for the average condition in this study. The calculation using the data of a typical day is as follows:

$$\begin{array}{ll}\text{change of film biomass} & = 31.5 \text{ mg/l} - 29 \text{mg/l} - (2.4 \text{ mg/l} \times 0.31)\\ (7/18/1978) & = 1.75 \text{ mg/l (sloughing)}\end{array}$$

Since the hydraulic loading on the same day was 0.147 m^3/m^2-day and there were 163.7 g of biomass in the biofilm,

$$\text{Rate of sloughing} = 1.75 \text{ mg/l} \times 0.147 \frac{m^3}{m^2\text{-day}} \times 23.3 \text{ m}^2$$

$$= 5.95 \text{ g/day.}$$

$$\text{Sludge retention time} = 163.7 \text{ g}/(5.95 \text{ g/day}) = 27.5 \text{ days}$$

The average sludge retention time for the biofilm (163.7 g biofilm in 125 liters of sewage in the RBC tankage, equivalent to 1310 mg/l MLSS) was 20.2-day in this study. The existing MLSS concentration in the RBC tankage was on the average 24 mg/l with a sludge retention time of 0.037 day (identical to the average hydraulic retention time in the RBC unit). The combined retention was therefore:

$$\text{Average RBC sludge retention time} = \frac{(1310 \times 20.2) + (24 \times 0.037)}{1310 + 24} = 19.8 \text{ days}$$

By comparison, the 19.8-day sludge retention time was longer than that of activated sludge processes and shorter than that of aerated lagoon processes. Since biological solid synthesis is inversely proportional to sludge retention time, it is then expected that sludge systhesis in the RBC unit should be between 0.5 (for activated sludge processes) and 0.15 (for aerated lagoons). This is in agreement with the synthesis value of 0.31 found in this study.

DESIGN EXAMPLE

Data developed from this study are used to design a retrofit RBC system to upgrade trickling filter effluents. Design criteria are listed below.

(a) Acceptable secondary effluent standards:

15 mg/l suspended solids
15 mg/l soluble BOD_5

Acceptable tertiary effluent standards:

1.0 mg/l suspended solids (dual media filter)
9.0 mg/l soluble BOD_5
2.0 mg/l soluble NH_3-N

(b) Effluent quality standards to be maintained throughout the year without any special allowance for cold weather conditions.

(c) High density media up to 132,000 ft^2 surface area for a 12-ft dia. module, 25-ft long shaft, and 4 stages.

(d) Use 20.4 m^3/m^2-day (500 gpd/ft^2) overflow rate for RBC clarifier design.

(e) Figure 3 alone for secondary treatment design, while Figures 3 and 5 both for tertiary treatment design.

(f) Trickling filter effluent and loading:

Design flow 1.0 mgd (3785 m^3/day)	
$SBOD_5$ concentration	22 mg/l
$SBOD_5$ loading	83.24 kg/day (183.5 lb/day)
Soluble NH_3-N conc.	8 mg/l
Soluble NH_3-N loading	30.26 kg/day (66.7 lb/day)

	Disc surface	Scale up Factor	No. of RBC module 12 ft dia. 25 ft shaft 4-stage	Clarifier Surface Area	
	m^2			m^2	ft^2
Secondary treatment	40,584 (436,860 ft^2)	1.5	4	185.8	2000
Tertiary treatment	67,300 (724,300 ft^3) based on $SBOD_5$ loading	1.5	6	use dual media filter with or without final clarifiers	
	97,870 (1,053,480 ft^2) based on soluble NH_3-N loading	1.2	8		

If RBC is to upgrade trickling filter effluents to meet tertiary standards, nitrification controls the size selection rather than BOD_5 removal since the allowable soluble NH_3-N loading is much lower.

ACKNOWLEDGEMENT

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UPGRADING EXISTING WASTE TREATMENT FACILITIES UTILIZING THE BIO-SURF PROCESS

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and

Robert J. Hynek
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Introduction

A large number of treatment plants around the United States have been forced to upgrade their effluent quality due to either hydraulic or organic overloading, or the establishment of more stringent effluent requirements by the Federal Government.

Since 1973 Autotrol Corporation has been developing new and innovative methods of using BIO-SURF, a Rotating Biological Contactor process, to upgrade existing treatment plants with minimal capital expenditures and little increase in the operational costs of the existing facility.

This paper presents the R.B.C. process fundamentals and various techniques Autotrol has utilized to upgrade existing primary clarification, trickling filtration and activated sludge plants.

The Bio-Surf Process

The BIO-SURF process is a Rotating Biological Fixed-Film Reactor. It employs large diameter corrugated and flat polyethylene sheets thermally welded together to form a media network for biological development. These polyethylene units are placed on a square steel shaft and the shaft and media are placed in a tank until approximately 40% of the media surface is submerged below the wastewater level. Micro-organisms naturally present in the wastewater adhere and develop on the media until a biological mass, approximately 1/16" thick, is developed on the media. The media is continually rotated alternately exposing the biological mass to the atmosphere and then to the wastewater. During the exposure of the biological mass to the atmosphere, oxygen present in the atmosphere is absorbed into the biological cultures present. When the biology rotates below the wastewater surface, the micro-organisms present on the media will assimilate and remove the soluble carbonaceous and/or nitrogenous impurities present in the waste.

The Rotating Biological Contactor, the BIO-SURF system, has been employed widely in the United States over the past four years, and has proven itself to be a viable and economic method for secondary treatment and nitrification and more recently denitrification. The main advantage of the system are its relatively simple operation and its significantly reduced power costs.

In the past the media has been rotated by a mechanical drive mechanism, i.e. a motor and gear reducer. More recently, Autotrol Corporation has developed an Air Drive system. This equipment consists of a series of cups or vanes attached to the periphery of the BIO-SURF media. A supplemental air source, installed below the media, releases air at low pressure into the vanes. Through the buoyant force of the air captured in the vanes the media is rotated. Controlling the air flow will change the shaft rotational speeds, thus providing a variable speed drive system.

It has been determined that the Air Drive system, AERO-SURF, has a number of biological advantages over the mechanical drive system. This is due to:

1. Higher dissolved oxygen within the reactor liquid;
2. Thinner biological cultures and its associated thinner anaerobic layer;
3. Elimination of sulfur bacteria and corresponding lower removal kinetics; and
4. The combination of the above factors results in an increase in removal efficiency per unit area.

Description of Upgrading Techniques

In employing the rotating biological contactor technology for expanding or improving existing treatment plant efficiency or capacity, a myriad of applications have developed. These R.B.C. systems include series and parallel operation with existing suspended growth and fixed growth systems and direct placement of R.B.C. equipment within primary clarifier and activated sludge basins. Various flow schemes are described in Chart 1 and Figures 1-11.

CHART I

TYPE OF FACILITY	UPGRADING TECHNIQUE	MODE OF APPLICATION	EXISTING LOCATION
Primary Treatment Plant	Increase Degree of Carbonaceous Removal	1) R.B.C. After Primary & Prior to a New Secondary Clarifier (See Figure 1)	
		2) Placement of R.B.C. In Primary Tanks with Horizontal Baffle (See Figure 2 & 3)	- Edgewater, NJ
Biological Fixed Film Reactors	Increase Hydraulic Capacity of Existing Facility	1) Place R.B.C. in Parallel with Existing Fixed Film Reactors. (See Figure 4)	- North Huntingdon, PA
ie: Trickling Filters	Increase Degree of Carbonaceous Removal	2) Place R.B.C. in Series with Existing Fixed Film (See Figure 5)	- North Huntingdon, PA
	Increase Treatment to Nitrification Capacity	3) Place R.B.C. in Series with Existing Fixed Film Reactor (See Figure 6)	- Birdsboro, PA
Biological Suspended Growth Reactors	Increase Degree of Carbonaceous Removal	1) Place R.B.C. prior to or in Parallel with Activated Sludge Reactor (See Figure 7 & 10)	
ie. Activated Sludge & its various modifications	Increase Treatment to Nitrification Capacity	2) Place R.B.C. in Series with Activated Sludge Process Train (See Figure 9)	- Cadillac, MI - Sarasota, FL
	Increase Hydraulic Capacity of Existing Facility	3) Place R.B.C. in Parallel with Activated Sludge Reactor (See Figure 10)	
	Increase Hydralic Capacity, Degree of Carbonaceous Removal, or Achieve Nitrification	4) Surfact - Placement of R.B.C. in Activated Sludge Reactor (See Figure 11)	- Philadelphia, PA

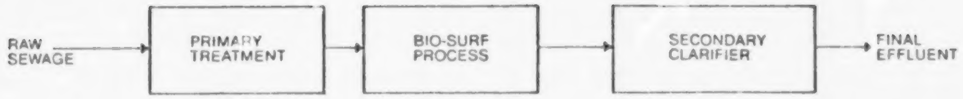

RAW
SEWAGE
PRIMARY
TREATMENT
BIO-SURF
PROCESS
SECONDARY
CLARIFIER
FINAL
EFFLUENT

Figure 1

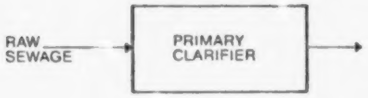

RAW
SEWAGE
PRIMARY
CLARIFIER

Figure 2

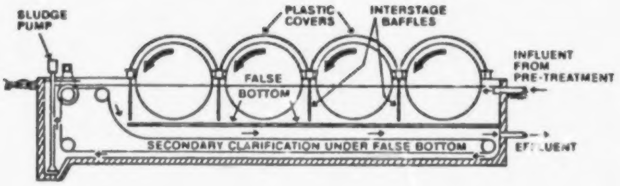
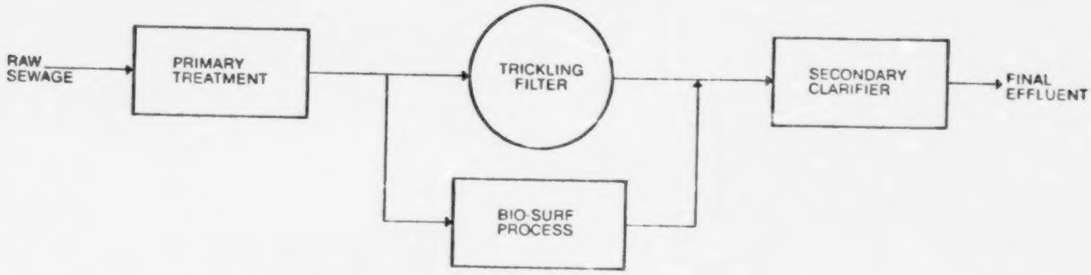

SLUDGE
PUMP
PLASTIC
COVERS
INTERSTAGE
BAFFLES
INFLUENT
FROM
PRE-TREATMENT
FALSE
BOTTOM
SECONDARY CLARIFICATION UNDER FALSE BOTTOM
EFFLUENT

Figure 3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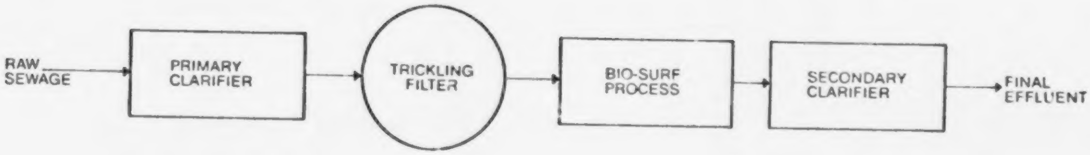
PRIMARY TREATMENT
TRICKLING FILTER
SECONDARY CLARIFIER
FINAL EFFLUENT
BIO-SURF PROCESS

Figure 4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PRIMARY CLARIFIER
TRICKLING FILTER
BIO-SURF PROCESS
SECONDARY CLARIFIER
FINAL EFFLUENT

Figure 5

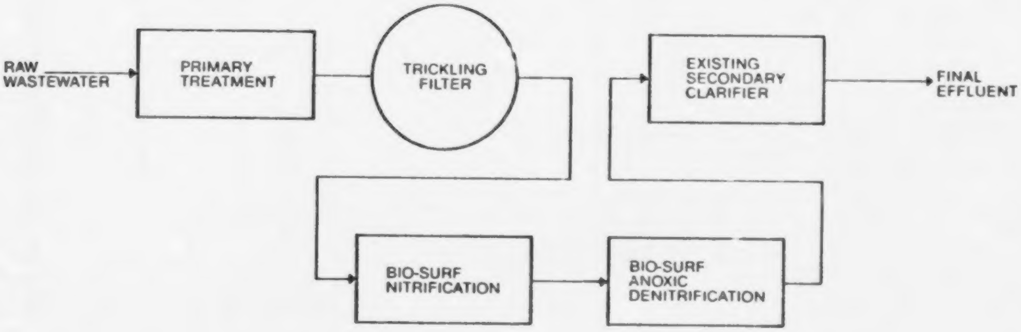

RAW WASTEWATER
PRIMARY TREATMENT
TRICKLING FILTER
EXISTING SECONDARY CLARIFIER
FINAL EFFLUENT
BIO-SURF NITRIFICATION
BIO-SURF ANOXIC DENITRIFICATION

Figure 6

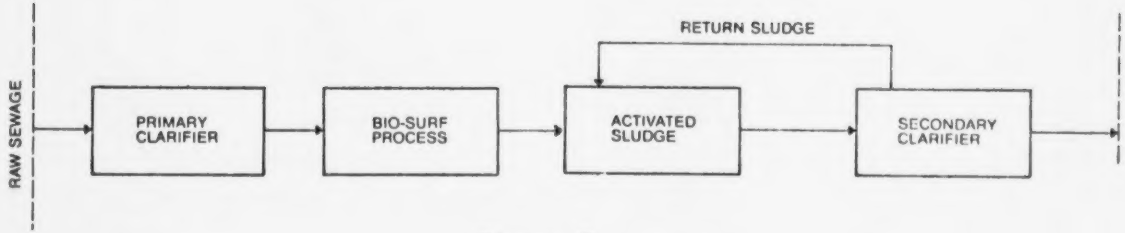
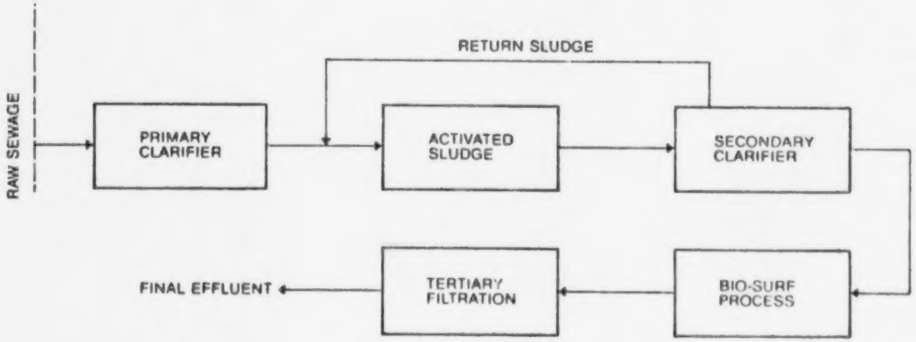
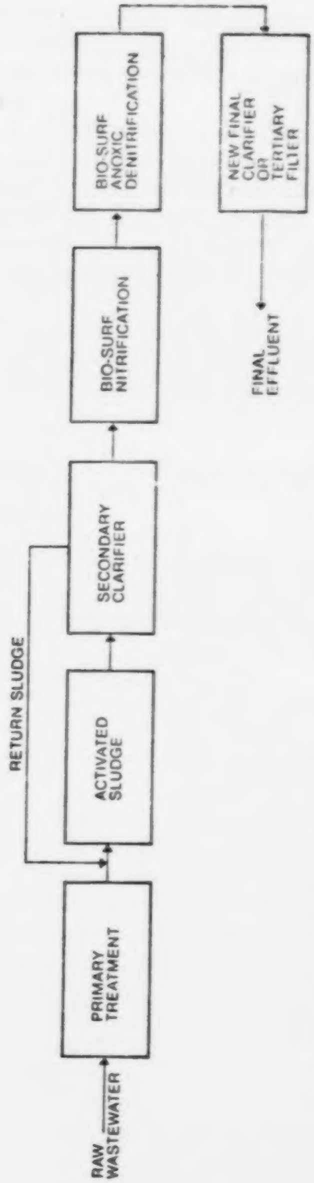

RAW SEWAGE
PRIMARY CLARIFIER
BIO-SURF PROCESS
RETURN SLUDGE
ACTIVATED SLUDGE
SECONDARY CLARIFIER

Figure 7


RAW SEWAGE
PRIMARY CLARIFIER
RETURN SLUDGE
ACTIVATED SLUDGE
SECONDARY CLARIFIER
FINAL EFFLUENT
TERTIARY FILTRATION
BIO-SURF PROCESS

Schematic flow diagrams for activated sludge plant upgrading

Figure 8


RAW WASTEWATER
PRIMARY TREATMENT
ACTIVATED SLUDGE
RETURN SLUDGE
SECONDARY CLARIFIER
BIO-SURF NITRIFICATION
BIO-SURF ANOXIC DENITRIFICATION
NEW FINAL CLARIFIER OR TERTIARY FILTER
FINAL EFFLUENT

Schematic flow diagram for activated sludge upgrading including nitrification/denitrification

Figure 9

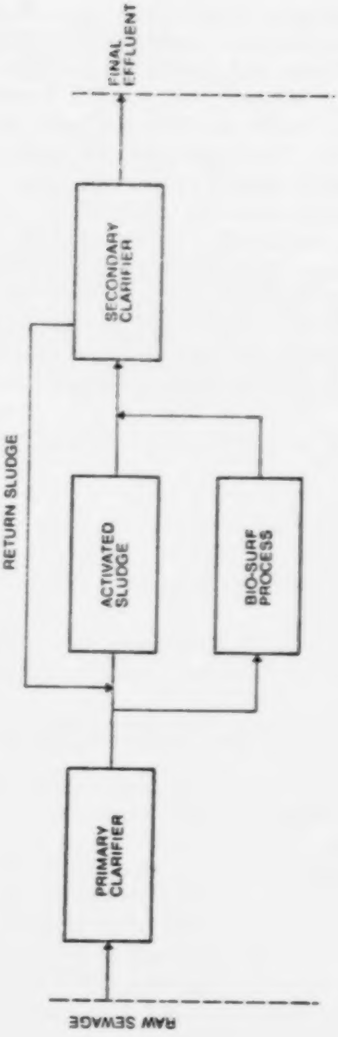

RAW SEWAGE
PRIMARY CLARIFIER
ACTIVATED SLUDGE
BIO-SURF PROCESS
RETURN SLUDGE
SECONDARY CLARIFIER
FINAL EFFLUENT

Figure 10

Application of the R.B.C. process to upgrading existing facilities is particularly attractive due to the following system features of the R.B.C. design.

1. Low hydraulic head loss - the head drop through R.B.C. tankage is less than 6" and therefore facilitates its utilization in existing treatment flow trains in various configurations.
2. Shallow excavations and tankage requirements - tankage is only 6' deep and therefore can be easily employed in locations where high groundwater and bedrock conditions are encountered. The shallow tankage requirements also facilitate employment within existing structures such as primary clarifiers and aeration basins.
3. Minimal effects on Solids Loading to secondary clarifiers - R.B.C. systems MLSS concentrations are extremely low and do not vary with seasonal fluctuations as compared to other attached growth systems or suspended growth systems. The impact of solids loading on downstream unit processes such as secondary clarifiers or final filters is minimal. Higher hydraulic rates and/or elimination of solid separation unit processes have proven feasible.
4. Modular Construction - Since the R.B.C. units are constructed in a variety of unit lengths and diameters which can be applied in a module-type approach, site-land availability presents minimal problems in employment of the R.B.C. concept.

Selection of a particular flow scheme from the various alternatives depends upon:

1. Site conditions (i.e.: land availability, hydraulic gradient of existing facility and soils conditions);
2. Sizing and maximum beneficial utilization of existing plant unit processes;
3. Economic consideration of tank modifications and/or construction;
4. Potential ease of future upgrading.

Upgrading Existing Primary Treatment Facilities

The concept of upgrading existing primary to secondary treatment without adding secondary clarifiers was first demonstrated by Autotrol Corporation at Edgewater, New Jersey. The Edgewater Sewage Treatment Plant has provided primary treatment as a mechanism for removing an average of 30% BOD and 52% suspended solids from a mixture of domestic and industrial wastewaters. The plant was built in 1954 and expanded in 1958 to a design capacity of 2.5 MGD. The plant site is extremely limited comprising only 1.5 acres of usable land area. The major original unit operations include pumping, comminution, grit collection, primary sedimentation and chlorination. Treated wastewater then flows to the Hudson River via an outfall discharge pipeline. Sludge is processed with two anaerobic sludge digesters and two vacuum filters. The treatment plant is currently processing its 1958 design capacity.

In 1972 the Interstate Sanitation Commission and the Environmental Protection Agency required that the plant meet secondary treatment standards, i.e. effluent monthly averages of 30 mg/l BOD and suspended solids, or a minimum of 85% BOD and suspended solids removal.

A concept for converting this plant to secondary treatment level was developed by Autotrol Corporation employing the BIO-SURF process. One of five existing primary treatment tanks, 70-feet long, 14-feet 4-inches wide, with an average side-water depth of 9-feet, was segregated and modified. This involved the division of the primary tank by an intermediate horizontal baffle; thus, the primary tank was divided into upper and lower chambers. The BIO-SURF units were placed in the upper zone to perform biological treatment of the influent wastewater, and original sludge collection mechanisms remained in the lower zone to collect suspended biological cultures which had sloughed from the R.B.C. units. Minimal primary treatment was employed prior to wastewater entry into the biological zone. The minimal primary treatment consists of grit removal and a much reduced primary clarifier zone with an overflow rate of 5,000 gpd/ft^2 at average flow. Startup occurred in 1974.

Performance for the year 1976 is shown in Table 1. This summary shows that the modified tank is currently treating 500,000 gpd with an average effluent of 29 mg/l BOD$_5$. This corresponds to an average effluent BOD and suspended solids removal of 89%.

Thus, by making minimal modifications to primary treatment facilities, the BIO-SURF process was able to convert an existing primary tank to a secondary treatment facility. The major modifications required to perform this function are:

1. The installation of a horizontal baffle, or false floor, separating the existing primary tank into separate zones;
2. The incorporation of Rotating Biological Contactors in the upper zone for removal of soluble carbonaceous or nitrogenous impurities in the wastewater. These Rotating Biological Contactors must be made to a specified length to conform to the width of the existing primary treatment tank;
3. Modifications to the sludge scraping mechanisms in the existing primary tank by re-directing the collector chain and putting in chain guides in the subjacent sedimentation zone.

Upgrading Existing Trickling Filter Plants

The BIO-SURF process has also been used widely in upgrading existing trickling filter plants. General flow diagrams for utilizing this concept would place the R.B.C. either in series operation with the trickling filter to upgrade the plant to meet more stringent effluent requirements, or in parallel operation with the trickling filter to upgrade the plant for increased flow or organic loading beyond the original design capabilities. (See Figures 4-7).

Presently in operation Autotrol Corporation has sewage treatment plants where the Rotating Biological Contactor system has been employed between the currently operating trickling filter and secondary clarifiers. Birdsboro, Pennsylvania, and North Huntingdon, Pennsylvania are two examples.

TABLE I

RESULTS OF PLANT SCALE TESTS - EDGEWATER, NEW JERSEY

PRIMARY TANK UPGRADING

MONTH	YEAR	FLOW	INF. BOD	INF. S.S.	EFF. BOD	EFF S.S.	% BOD	% S.S.
JAN	1976	0.560	128	106	24	23	81	78
FEB	"	N.R.	122	120	24	20	80	83
MAR	"	N.R.	179	152	31	24	83	84
APRIL	"	0.564	240	194	28	24	88	88
MAY	"	0.537	182	185	27	25	85	86
JUNE	"	0.480	283	272	26	20	91	93
JULY	"	0.540	325	336	33	22	90	94
AUG	"	0.395	300	266	39	53	87	80
SEPT	"	N.R.	367	313	27	20	93	94
OCT	"	0.446	324	237	34	23	90	90
NOV	"	0.433	283	231	21	18	93	92
DEC	"	0.400	302	310	33	26	89	89
AVG.		0.484	253	227	29	25	89	89

North Huntingdon, Pennsylvania, initially installed a rock trickling filter plant designed for 50% removal of BOD at a flow of 1.5 MGD. The entire plant was comprised of primary clarifiers, high rate rock trickling filters, secondary clarifiers, chlorine contact tanks, anaerobic digesters and sludge drying beds. Due to increased hydraulic flow and more stringent effluent requirements, it was necessary to increase the removal capability to 85% BOD removal at a flow of 1.75 MGD; thus the upgrade consisted of simultaneous increases in hydraulic capacity and treatment efficiency. The BIO-SURF system installed to meet the new conditions consists of four 20-foot long R.B.C. shafts providing 305,000 ft^2 of effective biological surface.

Although the R.B.C. currently operates in parallel with the rock filter system, the plant was designed to function either in series or in parallel operation with the original flow train. During new plant evaluations, it was observed that superior performance with regard to specific removal rate was attained when the system operated in the parallel mode (Figure 4). It is believed that the reason for better removals during the parallel operation mode was due to the nature of the substrate being more biodegradable in substance. Also, the higher substrate concentration increased the diffusion gradient to the microorganism. As a corsequence the plant was able to significantly upgrade the effluent quality to meet a total effluent BOD of 29 mg/l. Results of performance tests are shown in detail in Table II, for series operation and parallel operation.

The City of Birdsboro, Pennsylvania installation is similar to that of North Huntingdon, with the exception that the existing trickling filter was initially designed to provide greater percentages of BOD reduction. Results from the trickling filter showed effluent BOD and suspended solids of approximately 56 mg/l of BOD and suspended solids. The City of Birdsboro wished to upgrade their facility to meet effluents of approximately 25 mg/l Ultimate Oxygen Demand which required oxidation of both carbonaceous and nitrogenous impurities. Eight BIO-SURF units were installed between the existing trickling filter and secondary clarifier to upgrade the original system to meet the desired effluent quality. These units are currently operating at an overall hydraulic loading of 1.2 gpd/ft^2. Typical results are shown in Table III. In summary, then, data in Table II and Table III confirm the capability of the BIO-SURF process to upgrade plants where the trickling filter was initially designed as a roughing process, as at the North Huntingdon facility, or to provide nitrification to overloaded secondary plants, as at Birdsboro, Pennsylvania.

Upgrading Existing Activated Sludge Plants

1. Carbonaceous Reduction - As mentioned in the previous discussion, the R.B.C. process can be installed prior to existing suspended growth reactors to reduce the soluble BOD load to the suspended growth reactor. An example of such an installation is Crawfordsville, Indiana, which has a flow diagram similar to (Figure 7). Four shafts, which can operate as either single or multi-stage systems, follow the primary clarifier and precede the suspended growth reactor to reduce the influent soluble BOD concentration to the 25 to 50 mg/l range. This greatly reduces the soluble load to the suspended growth system and since the solids generated

TABLE II

SOLUBLE BOD$_5$ RESULTS - NORTH HUNTINGDON, PENNSYLVANIA

FOR RBC PERFORMANCE

HYDR.	INFL. BOD$_5$ to BIO-DISC			EFFL. BOD$_5$ FROM BIO-DISC		
LOAD (GPD/SF)	TOTAL (MG/L)	SOLUBLE (MG/L)	RATIO S/T (%)	TOTAL (MG/L)	SOLUBLE (MG/L)	RATIO S/T (%)
TEST B (SERIES, SUMMER)						
3.1	98	69	70	24	14	58
3.0	90	63	70	25	13	52
3.1	70	41	59	21	11	52
3.2	98	67	68	34	19	56
3.1	102	81	79	35	21	60
3.2	85	57	67	36	21	58
3.1	68	46	68	23	15	65
3.1	70	54	77	25	12	48
3.1	88	60	68	34	15	44
AVERAGE	85	60	70	29	16	55
TEST C (PARALLEL, SUMMER)						
3.3	140	86	61	26	20	77
3.1	123	87	71	--	--	--
3.1	119	86	72	25	12	48
3.7	121	87	72	23	12	52
3.0	118	87	74	31	15	48
4.0	121	79	65	34	20	59
6.5	60	30	50	22	11	50
3.2	140	99	71	32	14	44
5.2	107	67	63	36	21	58
AVERAGE	117	79	67	29	16	55

Autotrol Corporation
Reference No. 256-9

Table III

Laboratory Analysis Summary
Birdsboro, Pennsylvania
0.622 MGD, 1.13 gpd/ft^2, 19°C
October 11, 1979

		mg/l	
	P.C.E.	T.F.E.	S.C.E.
BOD_S	35	21	5
BOD_S (Carbon)	36	19	6
BOD_T	85	56	12
COD_S	80	70	35
COD_T	175	130	60
D.O.	-	-	-
HCO_3	125	110	55
N_K-Tot.	-	-	-
-Sol.	15.2	11.4	4.9
NH_3-N	8.5	7.3	1.6
NO_3-N	7.0	9.1	20.0
Org.N-Sol	6.7	4.1	3.0
-Insol.	-	-	-
pH	7.1	7.1	7.2
SS-Tot.	63	76	40
-Vol.	49	38	10
% VSS	(77.8)	(50.0)	(25.0)

by the R.B.C. system is relatively small, has little effect on solids inventory within the suspended growth reactor. Thus, by employment of the R.B.C. system in this manner the soluble load to the suspended growth reactor is significantly reduced allowing for increases in flow or degree of treatment. It is also anticipated that performance will demonstrate an over-all increase in stability against shock load conditions.

2. Nitrification - The BIO-SURF process has also been employed to upgrade existing activated sludge plants to provide nitrification. Operating experience at such plants include Cadillac, Michigan, Hinkley, Ohio and Sarasota, Florida.

   A flow diagram of the Sarasota, Florida installation is shown in Table IV. The BIO-SURF process is generally employed after the secondary clarifiers to provide additional BOD removal as well as nitrification. The system, because of its low solids generation during the nitrification stage, can discharge its effluent directly into the receiving water body or if higher degrees of suspended solids removal are required into tertiary filters. The latest data from this plant are shown in Table V. As can be seen by these results, extremely low BOD_5 and ammonia concentration have been obtained from the Sarasota, Florida installation.

SURFACT PROCESS

In 1976 a new concept for upgrading existing suspended growth reactors using Rotating Biological contactors was developed by the City of Philadelphia Water Department. This concept is called "The SURFACT Process" because it is a biological system resulting from the combination of rotating biological fixed film surfaces (SURF) and an activated sludge system (ACT). The purpose of merging the Rotating Biological Contactor with an activated sludge system is to provide additional removal of BOD or nitrification, and to provide process stability and flexibility. Studies since 1976 indicate also that the R.B.C. and the activated sludge system act, for the most part, independently of one another and BOD removals of 90% are possible in the same tankage where conventional activated sludge had obtained minimal treatment efficiencies.

The SURFACT concept involves only a slight modification of existing aeration tanks and installation of R.B.C. units in the upper regions of these tanks. By so doing, the Rotating Biological Contactors greatly increase the effective mixed liquor suspended solids concentration, (as much as 33% to 50%), and increases the removal capability in the aeration system without changing existing tank volumes, return sludge sizing, blower sizing, or any other aspect of the existing facilities. Also, since the equivalent increase in mixed liquor suspended solids is held on a fixed film reactor, the solids loading to the secondary clarifier is unchanged from the original design, thereby avoiding major modifications to the secondary clarifier. Rotation of an air drive R.B.C. can be provided by the roll of the existing aeration tank or by a supplemental air header placed below the media. Mechanical drive is another alternative.

As a result of preliminary testing conducted by the City of Philadelphia in 1974 and 1976, utilizing the AERO-SURF and the SURFACT approach, it was

Autotrol Corporation

TABLE V

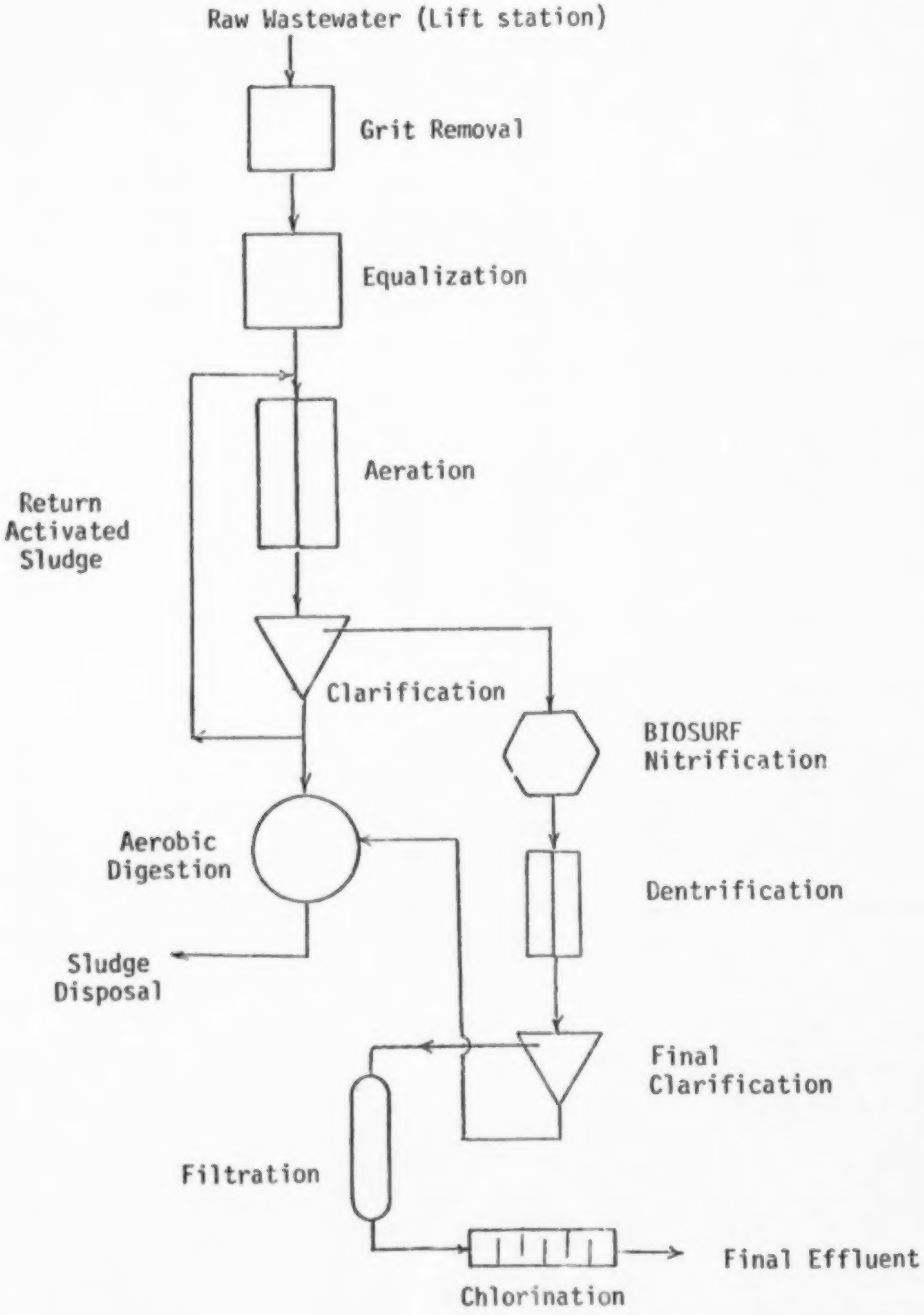
Gulf Gate, AWT, Sarasota, Florida

Year	Month	Flow MGD	BOD(mg/1)*		SS (mg/1)		Total N(mg/1)
			Infl.	Effl.	Infl.	Effl.	Effl.
1977	Jan	.864	177	1.3	278	1.1	1.4
	Feb	.860	198	2.7	279	1.5	2.3
	Mar	.853	195	2.2	274	1.6	2.0
	Apr	.792	190	1.0	288	1.7	1.0
	May	.760	198	0.7	310	1.0	1.4
	Jun	.783	183	0.5	282	1.0	1.7
	Jul	.786	217	0.6	338	0.5	1.8
	Aug	.900	207	0.7	245	0.6	2.1
	Sep	1.133	233	1.1	242	0.8	0.9
	Oct	.900	272	1.0	293	1.3	1.2
	Nov	.890	310	1.0	346	0.8	1.3
	Dec	1.000	251	1.4	251	0.2	0.6
1978	Jan	1.060	219	1.3	212	0.4	0.9
	Feb	1.120	242	1.9	220	1.1	3.3
	Mar	1.140	164	2.0	136	1.5	2.4
	Apr	.937	242	1.0	246	0.3	1.4
	May	.883	248	1.1	244	0.5	1.5
	Jun	.852	208	0.9	217	0.5	1.4
	Jul	.969	169	1.7	178	0.7	1.7
	Aug	1.080	153	1.2	196	0.6	1.7
	Sep	.922	189	0.7	204	0.5	1.1
	Oct	.905	191	0.9	212	0.4	1.2
	Nov	.926	210	1.1	243	0.4	1.2
	Dec	.980	199	0.9	205	0.4	1.4
1979	Jan	1.230	185	0.8	266	0.3	1.7
	Feb	1.170	208	0.9	273	0.6	1.4
	Mar	1.180	210	2.9	254	1.0	3.9
	Apr	1.080	231	1.5	282	0.5	2.0
	May	.990	208	0.8	299	0.4	1.3
	Jun	.973	201	0.6	278	0.6	1.2
	Jul	.974	190	0.7	230	0.8	1.1
Average		.964	210 **	1.2	254	0.8	1.6
		± .128	±33	±.6	±42	±0.4	±0.7

*Total

Table IV

GAC Utilities, Sarasota, Florida
Gulf Gate & South Gate STP


Raw Wastewater (Lift station)
Grit Removal
Equalization
Aeration
Return Activated Sludge
Clarification
BIOSURF Nitrification
Aerobic Digestion
Dentrification
Sludge Disposal
Final Clarification
Filtration
Final Effluent
Chlorination

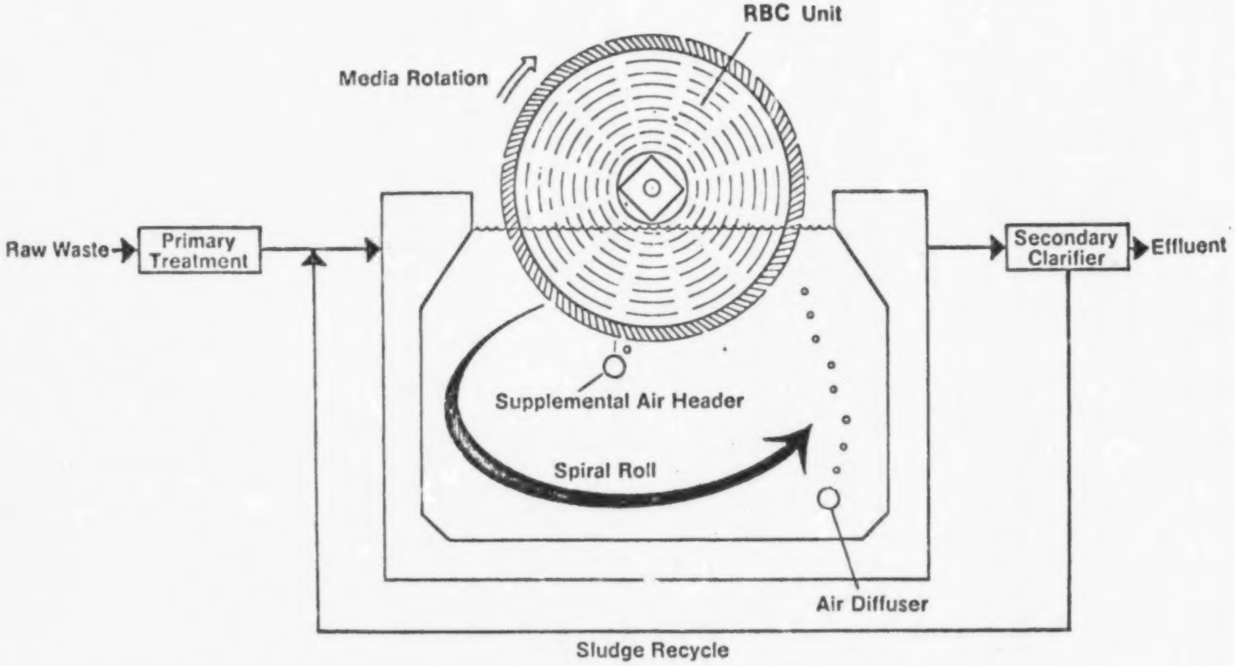
decided that a plant-scale SURFACT demonstration facility should be evaluated at the Philadelphia Northeast Plant. This treatment plant consists of grit removal, primary clarification, twenty existing aeration tanks and secondary clarifiers, and anaerobic digestion to process the resultant sludge from these unit operations. The plant-scale demonstration incorporated twenty-two AERO-SURF shafts into two parallel aeration basins. Each of these shafts contain media 12-foot in diameter with 100,000 square feet of effective surface area. The aeration basin was originally designed as a two-pass modified channel. Each pass was modified with eleven air-driven R.B.C. shafts. To receive the effluent from the SURFACT aeration tank a secondary clarifier at the existing facility was completely isolated by the installation of new return sludge pumps.

The evaluation at the Philadelphia plant began in October, 1977 and after mechanical start-up and process stabalization, collection of data was started shortly after January 1, 1978. The results of these tests are shown on Table VI.

It can be seen from Phase III and Phase IV of the operation that the SURFACT system has been able to consistently achieve BOD removals of 88% and 90% soluble BOD removal in excess of 95%, and suspended solids removal in excess of 80% at loadings of 74-pounds and 63-pounds of BOD per 1,000 cubic feet of aeration capacity. More recent evaluations demonstrate that loadings of 100# BOD_5/1000 cubic feet provided an effluent BOD quality of 20-25 mg/1 BOD. The F/M ratio necessary to achieve these results is approximately 0.5 with sludge age of approximately 2.7 to 3.6 days.

After 18 months of plant scale evaluation, it was concluded that the SURFACT Process was the most economical alternative for the upgrading of the Northeast Philadelphia facility. A total of approximately 180 shafts will be installed in the twenty passes.

In summary, the BIO-SURF, AERO-SURF and the SURFACT systems have proven complete flexibility in adapting to any existing plant's flow scheme in order to upgrade the facility. In addition, these processes have been able to perform this function in a consistently economical manner.


RBC Unit
Media Rotation
Raw Waste
Primary Treatment
Secondary Clarifier
Effluent
Supplemental Air Header
Spiral Roll
Air Diffuser
Sludge Recycle

The Surfact process consists of RBC *media assemblies installed in an activated sludge aeration tank.*

Figure 11

TABLE VI

SURFACT

	PHASE II	PHASE III	PHASE IV
% REMOVAL			
- SS	75	84	83
- BOD	82	88	90
- SBOD	91	95	95
EFFLUENT, MG/L			
- SS	25	26	23
- BOD	32	22	20
- SBOD	6	5	6
TOTAL % REMOVAL			
- SS	88	87	88
- BOD	83	90	90
- SOR, GPD/Ft.^{2}	720	720	600
- WOR, 10^3GPD/Ft.^{2}	82	82	68
DT, HOURS			
- PRIMARY	0.8	0.8	0.8
- AERATOR	3.0	3.2	3.6
- FINAL	2.1	2.2	2.4
FINAL, SOLIDS LOADING lbs/day/ft.^{2}	10	17	18
LBS BOD_A/1,000 Ft.^{3}	74	74	63
F/M			
- W/RBC	0.5	0.3	0.25
- W/O RBC	1.0	0.6	0.4
SRT			
-W/RBC	3.5	4.2	5.6
-W/O RBC	1.4	2.7	3.6
Ft.^{3}, AIR/Gal.	1.0	1.1	1.4
Ft.^{3}, AIR/1b BOD_A	600	700	840
LB O_{2A}/LB BOD_R	0.5	0.5	0.6
KWH/LB BOD_R	0.6	0.6	0.6

ROTATING BIOLOGICAL CONTACTOR PROCESS FOR SECONDARY TREATMENT AND NITRIFICATION FOLLOWING A TRICKLING FILTER

By

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Introduction

The practice of biological nitrification has been spurred by enactment of environmental Legislation as well as increased activities of regulatory agencies. Discharge limitations have been placed for ammonia-nitrogen in the National Pollutant Discharge Elimination System (NPDES) with Public Law 92-500 (currently the Clean Water Act of 1977, Public Law 95-217). Department of the Army installations must apply for and receive discharge permits for wastewater effluents. In many cases, existing NPDES permits impose ammonia-nitrogen limitations on Army-produced wastewaters, and it is anticipated that future permits will contain comparable, if not more stringent, discharge limitations.

Existing NPDES permits also impose secondary standards for biochemical oxygen demand of less than 30 mg/L, with the limiting concentration based on the conditions of the receiving waters. Biological treatment processes, which allow nitrification, decrease the BOD_5 level resulting from the carbonaceous and nitrogenous materials. As oxidation of carbonaceous organics and ammonia-nitrogen occur within the same process, stringent discharge requirements for both parameters are frequently obtained.

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The purpose of this study was to evaluate performance of the rotating biological contactor (RBC) for BOD_5 reduction and ammonia-nitrogen removal where the RBC followed a trickling filter. This treatment scheme was selected for evaluation based on several factors. It has potential utilization in existing facilities, a relatively low energy and operating cost, simplicity of operation, and flexibility with respect to expansion or upgrading of treatment facilities. The pilot RBC in this study received effluent from a high-rate trickling filter that performed at less than secondary effluent standards (i.e., trickling filter secondary effluent BOD_5 was routinely greater than 30 mg/L). Ammonia-nitrogen and BOD_5 removal efficiencies across the RBC were determined at varied hydraulic loading rates ranging from 1.5 to 4.3 gpd/sq. ft. of RBC surface area. Other parameters which were varied included the influent pH levels, organic loading rates, and suspended solids concentrations. Performance was evaluated during both summer and winter conditions.

Background

US Army wastewater treatment systems consist primarily of trickling filters as secondary treatment processes, a few activated sludge systems, and several extended aeration package plants. The Army has unique situations in which treatment plants are often flow underloaded due to decreases in the size of the Army population during peacetime. Also, consolidation of activities from several installations to only one installation, summer training of Reserve and National Guard troops, and maneuvers of troops can cause drastic seasonal changes in loadings on treatment plants. Drastic diurnal changes in loadings can be caused by civilian work forces that contribute wastes during normal working hours, but not at other times. In addition, size of Army treatment plants (0.1 to 5.0 MGD), dictate the need for simplicity of operation and maintenance. Therefore, advanced wastewater treatment (AWT) techniques applied to municipal wastewater treatment systems may not be applicable to Army treatment systems.

The RBC process was developed during the past two decades. In the last few years, the RBC process has been chosen over trickling filter and activated sludge systems in some designs. Reportedly, the RBC process allows for a longer and more intense contact time than the trickling filter process, thereby increasing BOD_5 removal. Continuous sloughing of biomass is a feature of an RBC process, whereas excess growth on a trickling filter can cause pondings. A high degree of of treatment within an RBC can be achieved by proper design of aeration capabilities and effluent retention without unnecessary recycling of effluent to maintain minimum wetting rates. The RBC process is less likely to be affected by organic shock loads or a hydraulic surge as would the activated sludge process, which relies on sludge recycling to maintain the proper food-to-microorganisms ratio. Lastly, because of the high density of the biological solids in the RBC process, purportedly higher flow rates can be used in the secondary clarifier, yet maintain a comparable or thicker sludge than the activated sludge process.

LITERATURE REVIEW

RBC Treatment Process

The RBC process consists of a series of plastic disks of which 40 percent of the surface area is immersed in wastewater (see Figure 1). As the disks rotate, the entire media surface develops a culture of microbiological organisms. The organisms adhere and multiply to form a uniform growth referred to as a fixed-film. The biomass supported by the plastic media picks up a thin layer of nutrient laden water as it rotates through the wastewater. The film of water trickles over the microorganisms which remove dissolved organics and oxygen. The rotation of the media through the wastewater not only allows for aeration and mixed liquor, but also provides shear forces which cause sloughing of excess growth.

RBC units commonly operate in series with the number of units depending on the organic and/or hydraulic load to be treated. The function of the first stages is to remove organic material, with subsequent stages removing ammonia when nitrification is used to meet effluent NH_3-N standards. Nitrification does not begin until the soluble-BOD_5 level and corresponding large populations of heterotrophic organisms have been adequately lowered. The actual reason that heterotrophs and autotrophic nitrifiers do not co-exist in equal quantities throughout successive RBC stages is not clearly understood, but it is reported that the activities of the two populations do not occur simultaneously.[1,2,3] The amount of nitrification achieved has been correlated to the hydraulic loading of the system, usually expressed as the volume of wastewater applied to a square measure of surface area per day. One to 4 gpd/sq. ft. have often been used as standard loading rates for pilot plants and full scale wastewater treatment facilities.[2,4,5] The change in hydraulic load also changes the organic load as more food is introduced to the active component of the waste treatment system.

Sufficient dissolved oxygen (DO) must be available in the wastewater within the RBC system to insure adequate treatment for BOD_5 removal and nitrification. Wastewater DO levels of 1 to 2 mg/L are generally considered to be the minimum requirement to avoid DO limiting conditions. Frequently, RBC systems have been designed to provide oxygen mass transfer via disk rotation through the wastewater and air. However, in some cases, this has been considered a shortcoming of the process since supplemental oxygen must sometimes be provided to prevent DO limiting conditions.[2]

Oxygen Demand of Wastewater

A major criterion used to determine the extent of pollution of receiving waters is the measurement of oxygen required for the stabilization of organic matter present in the system. The total amount of oxygen necessary to stabilize a waste is referred to as the oxygen demand. The ultimate oxygen demand includes not only the amount of oxygen required to stabilize oxidizable carbonaceous materials, but also that which is required to microbially transform ammonia-nitrogen to nitrate-nitrogen. For untreated domestic sewage there is little oxygen demand by nitrifier populations for the first 8 days of stabilization in the BOD test. Therefore, the BOD_5 test is normally considered as representing the oxygen demand of carbonaceous material. However, total BOD_5 can be poor indication of treatment where a significant population of nitrifying bacteria are present. For sewage that has received secondary treatment and nitrification, conversion of ammonia to nitrate in the BOD bottle may significantly increase the BOD_5 measurement and erroneously indicate a lesser degree of treatment than that actually received.

The purpose of biological treatment relative to carbonaceous material is the conversion of soluble organics to particulate bacteria.[7] However, the unfiltered BOD_5 test represents a measure of soluble as well as insoluble organic matter and NH_3-N oxidation. Biological treatment need not be applied to removal of colloidal and suspended organics. Suspended solids that contribute to oxygen demand can be removed by physical-chemical processes such as gravity settling and filtration. The practical consequence is that optimal treatment for removal of oxygen demand may be removal of suspended solids and not biological treatment. The use of filtered and unfiltered BOD_5 tests should indicate relative fractions of oxygen demand as originating from soluble or particulate material. In addition, the filtered BOD_5 test should not undergo nitrification, because initial nitrifying populations in the BOD bottle would be reduced to insignificant levels and the BOD bottle is subsequently seeded with raw sewage (i.e., heterotrophic bacteria). Thus, the unfiltered BOD_5 test can be an unreliable parameter from which to judge biological treatment performance.

Nitrogen Control

The principle of biologically induced nitrogen removal in wastewater treatment facilities is based on the activity of populations of autotrophic nitrifying and denitrifying bacteria and their capability to sequentially oxidize and reduce nitrogen from ammonia to nitrate to nitrogen gas. Nitrification is the oxidation of NH_3-N to nitrate, and denitrification is the reduction of nitrate to nitrogen gas. Different types of microorganisms are required for each action. The extent of their use in wastewater treatment depends upon the end objective. Nitrification is used to control wastewater effluent levels of ammonia, but both nitrification and denitrification must be used to control total nitrogen levels in wastewater effluents. Although process technology for ammonia-nitrogen removal includes breakpoint chlorination, ammonia stripping, ion exchange, and nitrification/denitrification, this paper deals only with nitrification.

In addition to nitrification/denitrification, microorganisms other than the nitrifiers and denitrifiers require nitrogen for growth. The amount of nitrogen assimilated during oxidation of carbonaceous material has been generally placed at 5 percent of the oxygen demand (i.e., BOD to N = 20 to 1).[8] The consequence is two fold: (1) nitrogen must be present for biological oxidation of carbonaceous material, and (2) removal of ammonia-nitrogen during biological treatment of wastewaters may be due to assimilation, not necessarily due to nitrification.

The importance of nitrogen control in wastewater effluents is its impact on receiving waters. As ammonia becomes oxidized to nitrate, the dissolved oxygen level of water is decreased. Ammonia-nitrogen at concentrations of 0.25 to 0.30 mg/L are lethal to fish within 14 to 21 days.[9] Nitrate is readily available for assimilation by plant life, causing algal blooms when present in too large a quantity.[16] Also, nitrate can cause methemoglobinemia in infants when contaminated water is used as a drinking water supply.[10]

Nitrification

The two microbial genera usually associated with nitrification are *Nitrosomonas* and *Nitrobacter*. Both genera of organisms are autotrophic nitrifying bacteria indicating that energy for growth is derived from the oxidation of inorganic nitrogen. The oxidation of ammonia to nitrate is a two step process requiring both organisms for the conversion. *Nitrosomonas* transforms ammonia to nitrite while *Nitrobacter* further oxidizes nitrite to nitrate. The overall oxidation of ammonia by these organisms is given by the following equation:

$$NH_4^+ + 2O_2 + 2\ HCO_3^- \xrightarrow[\underline{\text{Nitrobacter}}]{\underline{\text{Nitrosomonas}}} NO_3^- + 2H_2CO_3 + H_2O \qquad (1)$$

As ammonia is oxidized, carbonate is utilized, as nitrate formation occurs, carbonic acid is produced. This microbiologically induced change in the carbonate buffering system results in the destruction of alkalinity at a rate of 7.1 mg (as $CaCO_3$) per mg of ammonia oxidized.[11] As the nitrification process reduces the alkalinity and increases the carbonic acid concentration, the pH of the wastewater may drop as low as pH 6.0, and adversely impact the rate of nitrification. This decrease in pH can be minimized by aeration to strip CO_2 from the wastewater, or by insuring the presence of excess alkalinity.

Primary environmental conditions for optimizing rates of nitrification are pH and temperature. The reported pH optima cover a wide range, but the consensus is that as the pH decreases, the rate of nitrification declines. Sawyer, *et al.*,[12] and Engel and Alexander[13] have reported pH optima for nitrification between 8.0 and 9.0, and 7.0 and 9.0, respectively. Painter[14] has stated that nitrification processes cease at or below pH 6.3 to 6.7. Poduska and Andrews[15] have shown that abrupt changes in pH from 7.2 to 5.8 markedly reduced the ammonia oxidation by nitrifiers while the reversal in pH restored the original nitrification rate.

Temperature optima for nitrification are generally reported by various authors at about 30°C with a range of 28-35°C.[8,16,17,18,19,20,21] Temperature influences heterotrophic and autotrophic microorganisms, thereby affecting secondary treatment and nitrification efficiencies. The nitrification rate is more temperature sensitive than the rates for organic removal.[9] Nitrification rates decrease about 50 percent for each 10°C drop in wastewater temperature below about 30°C.[4] For example, the nitrification rate at 10°C would be about half that of 20°C. Secondary treatment efficiency is less likely to be affected by temperature changes, probably due to microbial population diversity and other system constraints. Organic removal rates for fixed-film processes should decrease about 25 percent for each 10°C drop in wastewater temperature below about 30°C.[9] For example, the rate of biological activity in a trickling filter process, at 10°C would be about 75 percent of that of 20°C. However, the actual temperature effect on a biological process is probably characteristic only of that system.

MATERIALS AND METHODS

The pilot RBC consisted of four compartments in series. The 0.5 meter plastic disks provided 250 sq. ft. of surface area for microbial attachment.

The disks are rotated through liquor at 13 rpm with 40 percent of the fixed-film submerged at any point. A schematic of the pilot RBC process is shown in Figure 1.

Pilot studies used domestic wastewater from the Fort Detrick housing area. The wastewater was shredded by a grinder pump and pumped into a 250 gallon equalization tank which was periodically replenished through a float level switch control. This enabled a relatively constant flow of partially settled wastewater to be pumped into the primary clarifier. The wastewater flowed by gravity through the primary clarifier and was then pumped to a high-rate trickling filter. The trickling filter contained a 2-inch irregular stone media with effective media depth of 4 ft. and 2.25 sq. ft. of filter surface. Effluent was collected at the bottom of the filter in a wet well and pumped to the secondary clarifier. This pretreatment scheme was used to simulate an existing trickling filter plant. The RBC followed trickling filter treatment and was used as an upgrading technique for additional BOD_5 reduction and nitrification (ammonia-nitrogen removal).

Performance of the RBC was evaluated over a two-year period without temperature control in order to determine both summer and winter efficiencies for BOD_5 reduction and ammonia-nitrogen removal. RBC performance was evaluated at normal and elevated pH levels of pH 7.1 and pH 8.7 for the RBC influent (trickling filter secondary effluent).

The RBC influent flow was varied to provide hydraulic loading ranging from 1.5 to 4.3 gpd/sq. ft. of RBC surface area. Treatment efficiency of the trickling filter was varied to provide different levels of BOD_5 in the RBC influent. Levels of suspended solids in the RBC influent were varied by having secondary settling before the RBC and having no settling between the trickling filter and the RBC.

Sampling and Analyses

Sample points included 24-hour composites of primary influent, trickling filter effluent (RBC influent) and RBC effluent. Analyses of test parameters in the four stages of the RBC unit were determined from grab samples.

Measurements of flow, temperature, dissolved oxygen, pH, suspended solids, total organic carbon, and ammonia-nitrogen were made daily. Alkalinity and chemical oxygen demand (COD) measurements were made four times per week. Biochemical oxygen demand (BOD_5) analyses were performed twice weekly.

Total organic carbon measurements were made on a Beckman Model 915 Total Organic Carbon Analyzer. Ammonia-nitrogen concentrations were measured with an Orion specific ion electrode. Dissolved oxygen and BOD_5 determinations were made using a Delta Scientific Model 2110 Dissolved Oxygen Meter and probe. Chemical oxygen demand, total Kjeldahl nitrogen, and phosphorus analyses were made using a Technicon Auto Analyzer II system according to Technicon methods. Filtered samples were filtered through fiberglass filters. All other analyses were performed according to Standard Methods.[22]

RESULTS AND DISCUSSION

Hydraulic Loading

The RBC was initially evaluated for secondary treatment and nitrification following a high-rate trickling filter during the fall, winter, and spring of 1976-1977 at hydraulic loadings of 1.5, 2.0, 3.0, and 4.0 gpd/sq. ft. of RBC surface area. Secondary treatment efficiency was evaluated using filtered and unfiltered BOD_5, and filtered TOC. Nitrification efficiency was evaluated using NH_3-N, TKN, NO_2/NO_3-N, and alkalinity. Suspended solids, pH, temperature, and dissolved oxygen levels within the RBC stages were monitored. Sample points included raw wastewater, trickling filter effluent and RBC effluent. Limited sampling was conducted within stages of the RBC to evaluate progression of treatment.

Secondary treatment performance of the RBC is shown by data in Table 1 in which the RBC removed 20 mg/L of TOC. Table 1 also shows environmental conditions of the study at 1.5 gpd/sq. ft. The wastewater temperature averaged 13.4°C during this period in January and February of 1977, and the median pH for RBC influent was 8.6. Natural (biological) recarbonation and nitrification within the RBC process depressed the pH to about 7.5 in the RBC secondary effluent.

Careful consideration should be used in evaluating treatment efficiency of the biological processes based on unfiltered BOD_5. The oxygen demand of suspended solids makes unfiltered BOD_5 levels higher than filtered BOD_5 levels for raw wastewater and trickling filter secondary effluent. For highly treated RBC secondary effluent, nitrification could add to the oxygen demand of suspended solids to make unfiltered BOD_5 greater than filtered BOD_5. Because of the effects of suspended solids and nitrification on oxygen demand, unfiltered BOD_5 should not be used alone to evaluate biological treatment efficiency. The difference between values of unfiltered and filtered BOD_5 must be recognized because NPDES permits only require analysis for unfiltered BOD_5 without distinguishing the soluble loading from the particulate loading of a biological treatment process. The difference between unfiltered and filtered BOD_5 values indicate that pilot studies did not optimize removal of suspended solids. This fact is also shown by the suspended solids data in Table 1. Data for ammonia-nitrogen in Table 1 indicate that nitrification was essentially complete for the RBC secondary effluent and could contribute little to the unfiltered BOD_5. Therefore, reductions in unfiltered BOD_5 values should be attempted by removal of suspended solids and not through biological treatment.

Figure 2 shows ammonia removal through biological activity in the trickling filter and RBC processes, where the RBC hydraulic loading was 1.5 gpd/sq. ft. The RBC secondary effluent ammonia level of 0.5 mg/L indicates this to be a highly nitrified wastewater. As discussed in the Literature Review section, removal of nitrogen can be due to both bacteriological assimilation and nitrification. Generally, nitrogen would be assimilated in the amount of about 5 percent of the quantity of carbonaceous oxygen demand utilized, or about 5 percent of the change in filtered BOD_5 across the treatment process. This nitrogen loss/removal should show up in the total Kjeldahl nitrogen (TKN) test (e.g., organic nitrogen plus ammonia) but may not decrease the ammonia level because microorganisms could assimilate nitrogen directly from organic compounds or convert organic nitrogen to ammonia. Removal of nitrogen beyond assimilation should be due to nitrification. The amount of nitrification can be determined from NO_2-N/NO_3-N analyses, since ammonia is oxidized to nitrate. Nitrite and nitrite-nitrate levels are shown in Table 1. The alkalinity test can also be used as a check on the amount of nitrification, because 7.1 mg/L of alkalinity are destroyed per 1 mg/L of NH_3-N oxidized.[11] Alkalinity levels are shown in Table 1.

Figure 2 shows the RBC influent and RBC secondary effluent levels of ammonia to be 18 and 0.5 mg/L, respectively. This shows that 17.5 mg/L of ammonia-nitrogen were removed across the RBC process, from both assimilation and nitrification. Unfortunately, TKN analyses were not available for this particular study because of equipment problems, and the total nitrogen story cannot be told here. Table 1 shows that 41 mg/L of filtered BOD_5 were utilized within the RBC process, therefore, 5 percent of 41, or about 2.0 mg/L, of nitrogen would be assimilated by heterotrophic organisms; additional nitrogen would be assimilated by autotrophic organisms. Remaining ammonia would be available for nitrification. Table 1 shows that 14 mg/L of NO_2-N/NO_3-N was formed and represents ammonia oxidized by nitrification. Table 1 also shows 100 mg/L of alkalinity destroyed across the RBC process, and this corresponds to 15 mg/L of ammonia nitrified (e.g., 100 divided by 7.1). In summary, of the nitrogen present in RBC influent, an excess of 2.0 mg/L was assimilated and about 15 mg/L was oxidized. This agrees well with the actual levels of ammonia observed. Ammonia removal was essentially complete in RBC secondary effluent at 1.5 gpd/sq. ft. even though temperature (13.4°C) was poor for nitrification; however, pH had remained near optimum for nitrification. The median pH levels were 7.5 in RBC secondary effluent.

Figure 3 contains results of further studies of secondary treatment by the RBC process following a trickling filter. Supporting data are contained in Table 1 for hydraulic loadings of 2.0, 3.0, and 4.0 gpd/sq. ft., respectively. Secondary standards were achieved at all hydraulic loadings as shown by filtered BOD_5 levels for RBC secondary effluent at all hydraulic loadings. Filtered BOD_5 utilized by biological activity within the RBC process (e.g., RBC influent values minus RBC secondary effluent values) were 36, 39 and 38 mg/L at 2.0, 3.0 and 4.0 gpd/sq. ft., respectively. This corresponds to an organic removal rate of about 1.3 lbs. filtered BOD_5/1000 sq. ft.-day at the 4.0 gpd/sq. ft. hydraulic loadings. The organic removal rates should increase if the organic loading were higher. Likewise, filtered TOC utilized within the RBC process were 21, 17, and 17 mg/L at 2.0, 3.0 and 4.0 gpd/sq. ft., respectively. The ratio of filtered BOD_5 utilized to filtered TOC utilized averaged 2.0:1. Wastewater temperature increased slightly at each increased hydraulic loading as shown in Table 1. RBC influent pH levels remained at about pH 8.5. It can be concluded from Figure 3 that an existing trickling filter can be upgraded to secondary standards by use of an RBC process at hydraulic loadings up to 4.0 gpd/sq. ft. and at organic removal rates in excess of 1.3 lbs. filtered BOD_5/1000 sq. ft.-day.

Figure 2 shows RBC treatment performance for nitrification at hydraulic loadings of 1.5, 2.0, 3.0 and 4.0 gpd/sq. ft. RBC secondary effluent was highly nitrified at 1.5 gpd/sq. ft. and at 2.0 gpd/sq. ft. for temperatures above 13°C (Table 1). Nitrification was essentially complete at 3.0 and 4.0 gpd/sq. ft. Wastewater temperature was relatively poor for nitrification at 1.5 and 2.0

gpd/sq. ft. (e.g., 13.4 and 15.1°C, respectively) as shown by data in Table 1. However, pH levels were near optimum for nitrification at those loadings. RBC secondary effluent pH was about pH 7.5. Wastewater temperature and pH levels were relatively good for nitrification at hydraulic loadings of 3.0 and 4.0 gpd/sq. ft. as shown by data in Table 1. Continued good performance by the RBC process at the higher hydraulic loadings is probably attributable to environmental conditions of wastewater temperature and pH.

Combined use of Figures 2 and 3 and Table 1 with supporting data are necessary to fully evaluate nitrification across the RBC process. Figure 3 showed that about 40 mg/L of filtered BOD_5 was utilized across the RBC process at each hydraulic loading; this represents about 5 percent of 40 or 2.0 mg/L of nitrogen assimilated by heterotrophic organisms. Additional nitrogen would be assimilated by autotrophic nitrifying organisms. Data in Table 1 show nitrate levels present in RBC secondary effluent, and show the alkalinity destroyed by the RBC process to compare favorably with ammonia levels. For example, at the hydraulic loading of 4.0 gpd/sq. ft., 63 mg/L of alkalinity were destroyed, corresponding to 8.9 mg/L of ammonia-nitrogen removal by nitrification. To balance nitrogen losses at 4.0 gpd/sq. ft., 8.9 mg/L would be added to 2.0 or more mg/L, showing total nitrogen assimilated and nitrified to be about 11.0 mg/L plus 2.0 mg/L in RBC secondary effluent, or 10 mg/L ammonia-nitrogen removed at 4.0 gpd/sq. ft. Organic nitrogen was unaccounted for and would represent a fraction of the nitrogen assimilated or present as ammonia in the RBC secondary effluent.

The effect of wastewater temperature on secondary treatment and nitrification across the RBC process can be seen by use of Figure 4. This Figure shows the temperature profile and RBC secondary effluent ammonia levels during the period December 15, 1976, to January 11, 1977, at a hydraulic loading of 2.0 gpd/sq. ft. The RBC secondary effluent was highly nitrified during the period of December 15-28 as evidenced by an average ammonia-nitrogen level of 1.0 mg/L. Ammonia nitrogen levels in the RBC secondary effluent increased to about 6.0 mg/L during the period of January 3-11, after wastewater temperatures had dropped to about 10°C. Decreased ammonia removal lagged behind this temperature drop, but the effect was substantial in lessening the degree of nitrification. Both organic removal and nitrification rates should have been affected by the drop in wastewater temperature. Decreased nitrification rates should be about half for each 10°C drop,[11] while organic removal rates (secondary treatment) should be about 75 percent for each 10°C drop.[7] Combined temperature effects from decreased organic removal and nitrification rates resulted in the increased ammonia levels shown in Figure 4. Treatment progression within the RBC process should show those effects, since organic removal occurs in the initial stages and nitrification in the later stages.

Table 2 shows the progression of nitrification within the RBC process at the 1.5 gpd/sq. ft. hydraulic loading. Grab samples were used for data presented in Table 2 and should represent a valid picture of RBC stage-by-stage performance, because flow equalization had occurred within treatment processes before RBC treatment. Nitrification was occurring in Stage 1 and subsequent stages as evidenced by formation of nitrate and removal of ammonia. It is also evident that nitrification and organic removal were occurring simultaneously within the initial stages because the filtered BOD_5 and TOC levels dropped across the RBC process (Table 1), even though organic removal data was not taken within RBC stages.

Ammonia was removed in greater quantity than nitrate formed; the difference, about 3.0 mg/L, is attributed to assimilation by both heterotrophs and autotrophs. Dissolved oxygen levels remained ideal for nitrification in all RBC stages. The pH drop from stage to stage was caused by biological recarbonation plus destruction of alkalinity by nitrification.

Figure 5 shows progression of nitrification within the RBC process at various hydraulic loadings of 1.5, 2.0, 3.0 and 4.0 gpd/sq. ft. Rates of nitrification appeared to reach a maximum level at each loading, and then decreased in the last RBC stage. This points out several issues of ammonia-nitrogen removal. First, nitrification rate is influenced by organics removal, and organic removal seems to take precedence. Ammonia removal in Stage 1 shows this influence by organic removal. Ammonia removal in Stages 2 and 3 indicate that nitrification rates are relatively unaffected by ammonia concentrations above about 5 mg/L, while State 4 results indicate that nitrification rate has decreased due to limited concentrations of ammonia. That is, nitrification rates approximate a zero-order reaction above about 2.5 mg/L and first-order reaction below 2.5 mg/L.[11] This explains why stringent NPDES permit limitations of $\leq$2.0 mg/L makes compliance difficult.

No Intermediate Settling

Further studies of the RBC process for secondary treatment and nitrification following a trickling filter were made at hydraulic loading of about 3.0 gpd/sq. ft. without intermediate settling between the trickling filter and RBC process. These studies were conducted to evaluate the possibility of upgrading existing trickling filter plants by use of RBC processes while maintaining existing secondary clarifiers to settle out RBC effluent. This scheme should result in savings on total wastewater treatment plant upgrade by making maximum use of existing facilities. These studies were conducted with RBC influent pH elevated to about pH 8.0 to create good environmental conditions for nitrification, and separately with neutral pH for the RBC influent to evaluate pH effect on nitrification.

Table 3 contains data for secondary treatment performance of the trickling filter and RBC treatment processes at 3.0 gpd/sq. ft. without intermediate settling between the trickling filter and RBC process. Filtered BOD_5 values show that the RBC secondary effluent met secondary standards. It can be concluded from data in Table 3 that the RBC process can be used to upgrade an existing trickling filter to secondary standards without intermediate settling between the trickling filter and RBC process.

Progression of treatment within the RBC stages at 3.0 gpd/sq. ft. following a trickling filter without intermediate settling is shown in Figure 6. This shows that rates of nitrification were relatively low in Stage 1 and higher in Stages 2 and 3, and again lower in Stage 4. This is similar to results of Figure 5 where intermediate settling had been used before the RBC process. Again, these rates of nitrification show that organic removal affects nitrification rates in the initial RBC stages where secondary treatment takes place and nitrification rates are highest at ammonia levels above about 5 mg/L where the rate of reaction is reported to be zero-order.[11] Nitrification rates decrease when the ammonia level drops below about 5 mg/L so that the latter stage of treatment is less efficient, and stringent NPDES permits become more difficult and costly to meet. It can be concluded from use of Figures 5 and 6 that nitrification across an RBC process is not greatly influenced by intermediate settling between the trickling filter and RBC process.

The effect of pH on nitrification across the RBC process is shown in Figure 7, where the RBC process followed the trickling filter without intermediate settling. RBC secondary effluent ammonia levels increased from about 1.5 mg/L with elevated pH adjustment before the RBC process to about 10 mg/L at neutral pH levels. Good nitrification within the RBC process continued for about 4 days after elevated pH changed to neutral levels. Then, poor nitrification continued for the duration of the study at neutral pH. Figure 7 shows the benefit of chemical feed for pH and alkalinity control for nitrification. Reportedly, elevated pH primarily increases the ratio of nitrifiers to heterotrophs in the system.[11] This favors simultaneous nitrification and organic removal in initial RBC stages and higher nitrification rates in latter stages for elevated pH levels, but not for neutral pH.

Further study of the pH effect on nitrification was conducted by repeating the neutral pH conditions for RBC performance without intermediate settling. Data are presented in Table 3. Figure 8 shows ammonia levels for RBC influent and RBC secondary effluent. RBC performance for nitrification was relatively good with elevated pH as indicated by an effluent ammonia level of 1.6 mg/L, but poor with neutral pH as indicated by effluent ammonia levels of 12.7 mg/L and 8.8 mg/L for the two studies. Supporting data in Table 3 for alkalinity and TKN confirm the poor nitrification at neutral pH levels; 66 mg/L of alkalinity were destroyed indicating 9.3 mg/L of ammonia nitrified at neutral pH values.

Data in Table 3 for filtered BOD_5 and TOC indicate that secondary effluent standards were achieved by the RBC process at neutral pH. RBC secondary effluent filtered BOD_5 values were 8 and 10 mg/L for the two studies, and 21 and 22 mg/L of TOC were removed by the RBC process. This secondary treatment performance by the RBC process at both neutral pH levels compares favorably with RBC performance at elevated pH levels of about pH 8.0 shown by data in Table 3. It can be concluded that elevated pH levels of about pH 8.0 do not affect secondary treatment performance (BOD_5 removal) of the RBC process.

Organic Loading

Further studies of the RBC process for secondary treatment and nitrification were made at hydraulic loadings of about 3.0 gpd/sq. ft. to evaluate the effect of organic loading on secondary treatment and nitrification. These studies evaluated the effect of partial and complete secondary treatment by an existing trickling filter, versus no secondary treatment, prior to an RBC process for secondary treatment and/or nitrification. These studies should help answer the question arising in design upgrades, "Should existing facilities be abandoned, or used?" The result should be that increasing organic loading increases effluent ammonia levels unless additional RBC surface is available at higher organic loadings, because substantial organic removal must occur before nitrification begins.[1,11] Therefore, the use of existing facilities for organic removal should substantially reduce surface area requirements of the RBC process for further secondary treatment plus nitrification. For example, Antonie showed that a 6 MGD, mechanical drive RBC plant for secondary treatment to 20 mg/L BOD_5 would cost \$3 million (in 1976) while such a plant that provided both secondary treatment and nitrification to 1.0 mg/L ammonia would cost \$4.5 million.[5] This points out the value of existing facilities, such as trickling filters, that provide organic removal, during design upgrade. In the example cited, an existing trickling filter providing secondary treatment could be worth up to \$3 million.

Earlier studies at elevated pH levels using intermediate settling between the trickling filter and RBC process were based on hydraulic loadings. However, the organic and ammonia concentrations remained essentially constant, so that increases in hydraulic loading also increased organic loading. Results of these studies at elevated pH levels showed that increased hydraulic (and organic) loadings had little effect on effluent ammonia levels (Figure 4). However, temperature increased with increases in loadings during those studies and partially accounted for additional nitrification at higher organic loadings. In addition, it was observed that simultaneous organic removal and nitrification occurred in the same stages of the RBC process at elevated pH levels, and helped account for the good nitrification at high loadings.

Studies were conducted at neutral and elevated pH levels and two temperature values to further evaluate the organic loading effect on secondary treatment and effluent ammonia levels. The RBC process was evaluated at 3.0 gpd/sq. ft. following a trickling filter at wastewater temperatures of about $25^{\circ}C$ and $17^{\circ}C$ at pH levels of 7.1 and 8.5 respectively. Also, the RBC process was evaluated without prior biological treatment at wastewater temperatures of about $25^{\circ}C$ and pH 8.8. Results of RBC treatment performance for secondary treatment are shown in Figure 9, and nitrification results are shown in Figure 10. Supporting data are contained in Table 4.

Figure 9 shows that RBC secondary effluent met secondary standards in all cases. However, this does not imply that organic removal within the RBC process was the same in all cases.

Comparable filtered BOD removal was observed when the trickling filter preceded the RBC at 17 and $25^{\circ}C$. However, considerably more NH_3-N was removed at $17^{\circ}C$, not as a result of temperature, but because the influent pH had been raised to pH 8.5. When the trickling filter was not used, applying an increased organic load to the RBC, a greater reduction in filtered BOD occurred. However, the ammonia removed was decreased when the increased organic load was applied to the RBC.

Figure 10 shows that nitrification was near completion at $25^{\circ}C$ with lower organic loading on the RBC process; whereas, nitrification was relatively poor at the lower temperatures with higher organic loading to the RBC process. These results indicate higher organic loadings, which in turn increase effluent ammonia levels, because substantial organic removal must occur before nitrification begins. Lower temperatures seemed to have a similar effect on effluent ammonia levels, like that observed with increased organic loading.

SUMMARY AND CONCLUSIONS

The RBC process performed effectively as a system to upgrade an existing trickling filter plant when used for secondary treatment over a range of hydraulic loads from 1.5 to 4.3 gpd/sq. ft. Secondary treatment was less noticeably affected by wastewater temperature than was nitrification. Results also indi--cated that elevating the RBC influent pH to about pH 8.0 increased the nitrification rate. The increase in nitrification rate decreases the RBC surface area requirements, resulting in less costly wastewater treatment plant upgrades.

An existing trickling filter plant can be upgraded to secondary standards for BOD_5 by using a rotating biological contactor (RBC) process for organic removal without intermediate settling between the trickling filter and RBC process.

An existing trickling filter plant can be upgraded for ammonia removal by using an RBC process for nitrification without intermediate settling between the trickling filter and RBC process.

The use of existing trickling filter plants for partial or complete secondary treatment decreases the surface area requirements of RBC's for secondary treatment and/or nitrification, resulting in less costly upgradings.

Chemical feed, or elevation of pH levels to about pH 8.0 in RBC influent, increases the rate of nitrification and decreases RBC surface area requirements, resulting in less costly improvement and more consistent attainment of ammonia removal.

Wastewater temperature affects both secondary treatment and nitrification within the RBC process, with nitrification rates being more sensitive to low temperatures than organic removal.

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DISCLAIMER

"The opinions or assertions contained herein are the private views of the authors and are not to be construed as official or as reflecting the views of the Department of the Army or the Department of Defense."

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TABLE 1. RBC TREATMENT PERFORMANCE AT VARIED HYDRAULIC LOADING RATES (gpd/sq. ft.)

	1.5		2.0		3.0		4.0	
Parameter (mg/L)[a]	RBC Inf	RBC Eff	RBC Inf	RBC Eff	RBC Inf	RBC Eff	RBC Inf	RBC Eff
BOD_5								
Filtered	50	9	45	9	64	9	49	11
Unfiltered	99	23	95	32	100	-	81	-
NH_3-N	15.9	0.7	15.2	1.5	12.2	2.7	12.5	2.8
TOC, Filtered	35	15	32	11	35	18	32	15
SS	102	33	101	25	112	72	55	26
Alkalinity	231	127	212	140	215	157	209	146
pH[b]	8.6	7.5	8.3	7.5	8.5	7.4	8.0	7.3
Temperature, °C	13.4	-	15.1	-	16.8	-	20.6	-
NO_2-N	0.18	0.22	0.1	0.5	0.18	0.39	-	-
NO_3-N	0.44	15.65	0.4	14.9	0.16	11.30	-	-

a. Mean values

b. Median value

TABLE 2. NITRIFICATION WITHIN THE RBC TREATMENT PROCESS AT VARIED HYDRAULIC LOADING RATES

RBC Stage Number	Parameter				
	NH_3-N	pH	DO	°C	$NO_2 + NO_3$-N
I. 1.5 gpd/sq. ft.					
1	9.4	8.0	4.6	13.2	4.68
2	3.6	7.5	5.1	12.7	9.60
3	1.8	7.4	5.9	12.4	14.56
4	0.9	7.5	7.2	12.2	17.24
II. 2.0 gpd/sq. ft.					
1	13.4	7.9	3.5	12.7	-
2.	8.4	7.7	4.1	12.2	-
3	3.7	7.6	4.6	11.9	-
4	1.8	7.5	5.0	11.7	-
III. 3.0 gpd/sq. ft.					
1	11.1	8.2	5.1	17.5	4.70
2	9.7	7.7	4.9	17.3	5.73
3	6.2	7.5	5.1	17.2	9.15
4	3.6	7.4	5.3	17.3	13.60
IV. 4.0 gpd/sq. ft.					
1	12.7	7.9	4.5	20.2	-
2	9.6	7.4	4.2	19.8	-
3	5.3	7.2	4.2	19.7	-
4	2.3	7.2	4.8	19.7	-

TABLE 3. RBC TREATMENT PERFORMANCE AT 3.0 gpd/sq. ft. WITHOUT INTERMEDIATE SETTLING

Parameter	(Summer 1977) RBC Inf	(Summer 1977) RBC Eff	(Spring 1978) RBC Inf	(Spring 1978) RBC Eff	(Late Summer 1977) RBC Inf	(Late Summer 1977) RBC Eff
BOD_5 (mg/L)						
Filtered	39	8	57	10	53	8
Unfiltered	146	ND	ND	ND	122	18
TOC (mg/L)	43	22	46	24	ND	ND
NH_3-N (mg/L)	19.5	8.8	24.6	12.7	18.2	1.6
TKN (mg/L)	27.1	10.4	24.3	ND	26.1	4.2
Alkalinity (mg/L)	204	140	167	101	266	153
pH[a]	7.0	7.1	6.7	6.9	7.7	7.2
Temperature, °C	24.9	ND	17.4	ND	23.9	ND

[a] Median value

TABLE 4. RBC TREATMENT PERFORMANCE AT 3.0 gpd/sq. ft. AND VARIED ORGANIC LOADS

Parameter	Trickling Filter plus RBC Process, 25°C		Trickling Filter plus RBC Process, 17°C		RBC Process, 25°C	
	RBC Influent	Effluent	RBC Influent	Effluent	RBC Influent	Effluent
BOD_5 (mg/L)						
Filtered	43	9	47	9	73	5
Unfiltered	136	22	90	ND	161	11
NH_3-N (mg/L)	19.5	8.8	12.9	3.4	24.4	12.6
TKN (mg/L)	27.1	11.6	ND	ND	28.8	13.5
COD, Filtered (mg/L)	98	38	83	36	197	134
Alkalinity	203	140	228	159	223	193
pH[a]	7.1	7.1	8.5	7.4	8.8	7.2
Temperature	24.2	25.5	17.5	17.0	24.4	ND

a. Median value

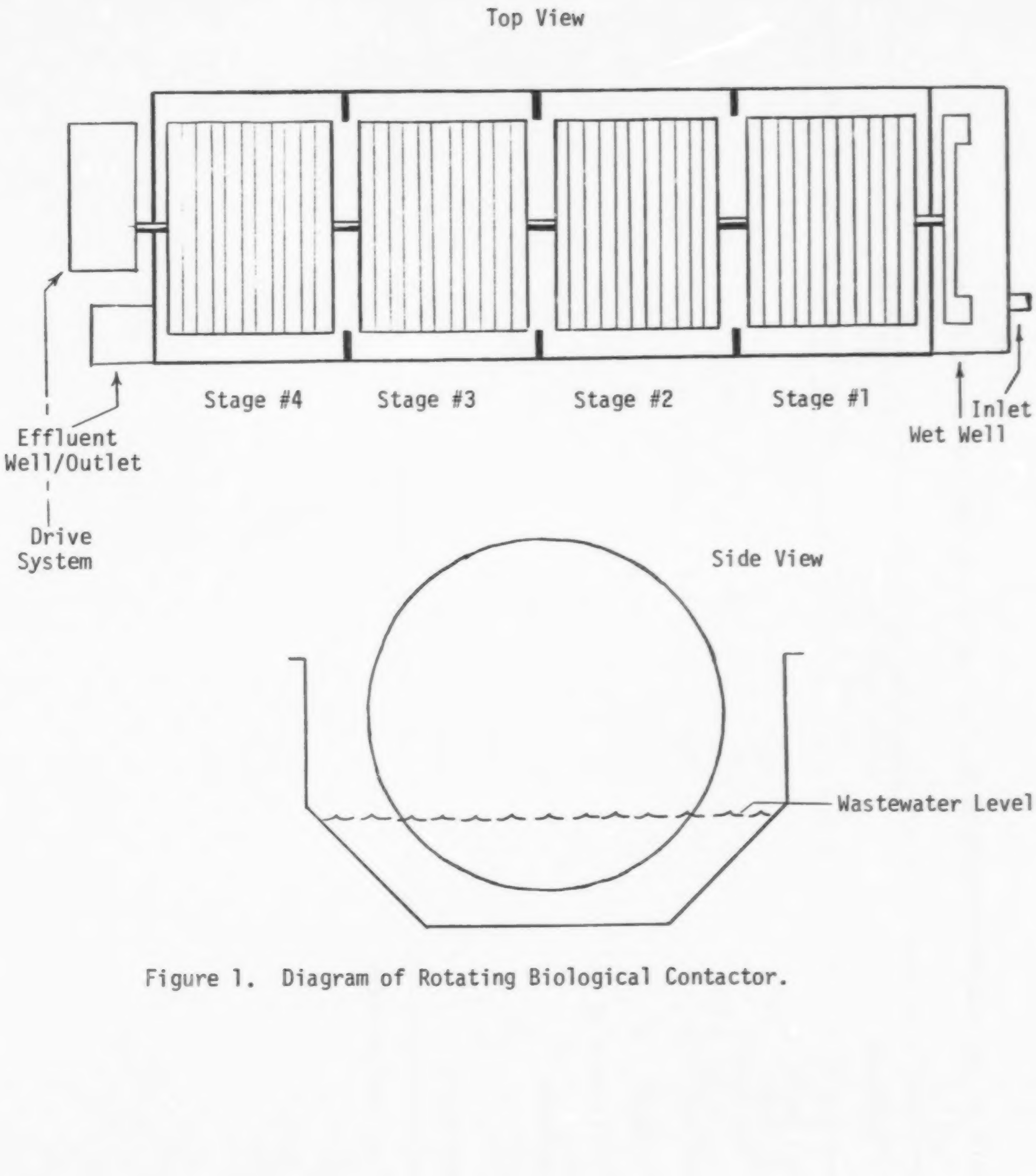
Top View
Stage #4
Stage #3
Stage #2
Stage #1
Inlet
Wet Well
Effluent Well/Outlet
Drive System
Side View
Wastewater Level

Figure 1. Diagram of Rotating Biological Contactor.

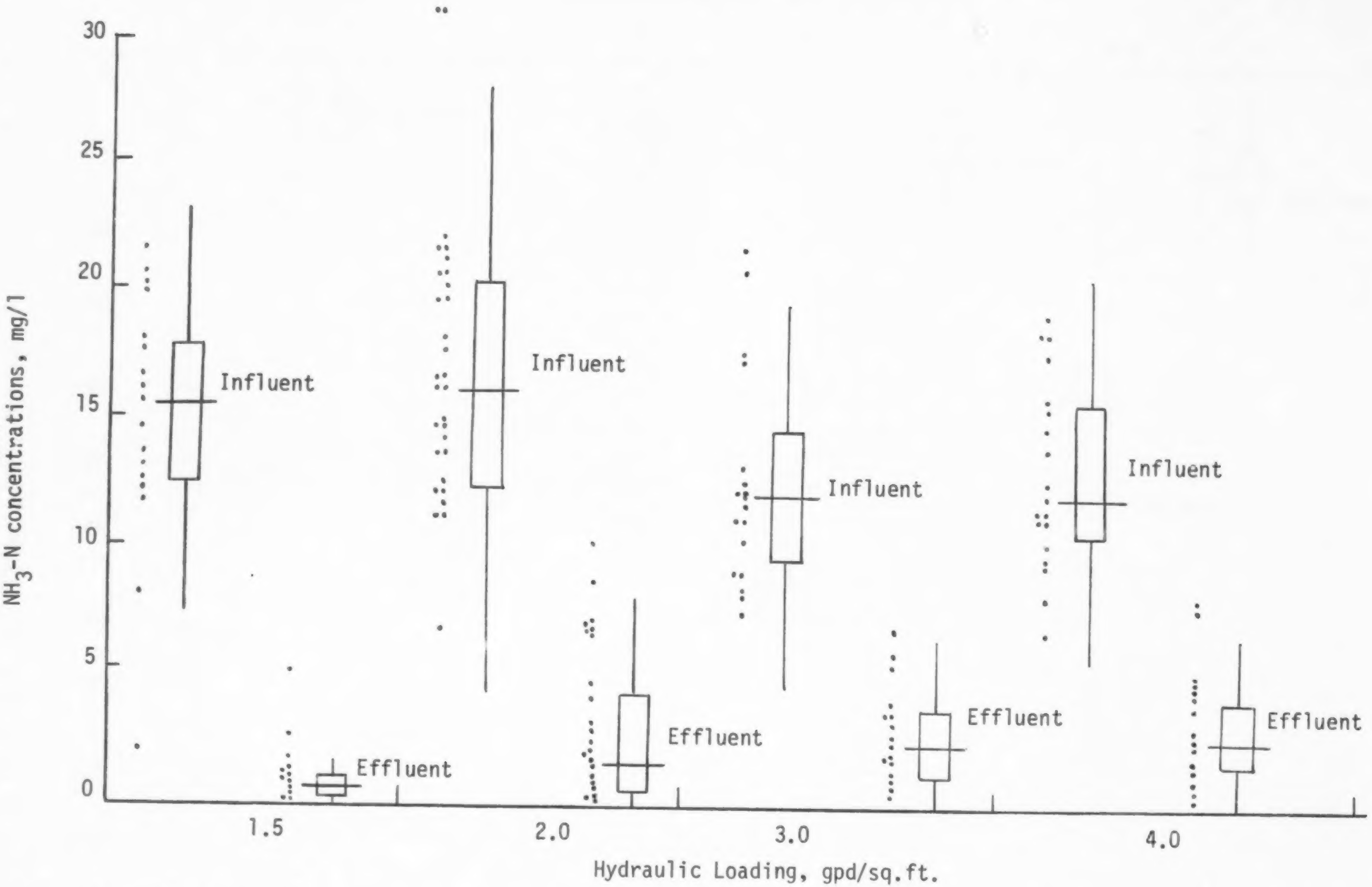

Figure 2. RBC Performance for Ammonia-Nitrogen at Various Hydraulic Loading Rates. RBC influent concentrations of ammonia-nitrogen are compared to RBC secondary effluent concentrations at each hydraulic loading rate, respectively.

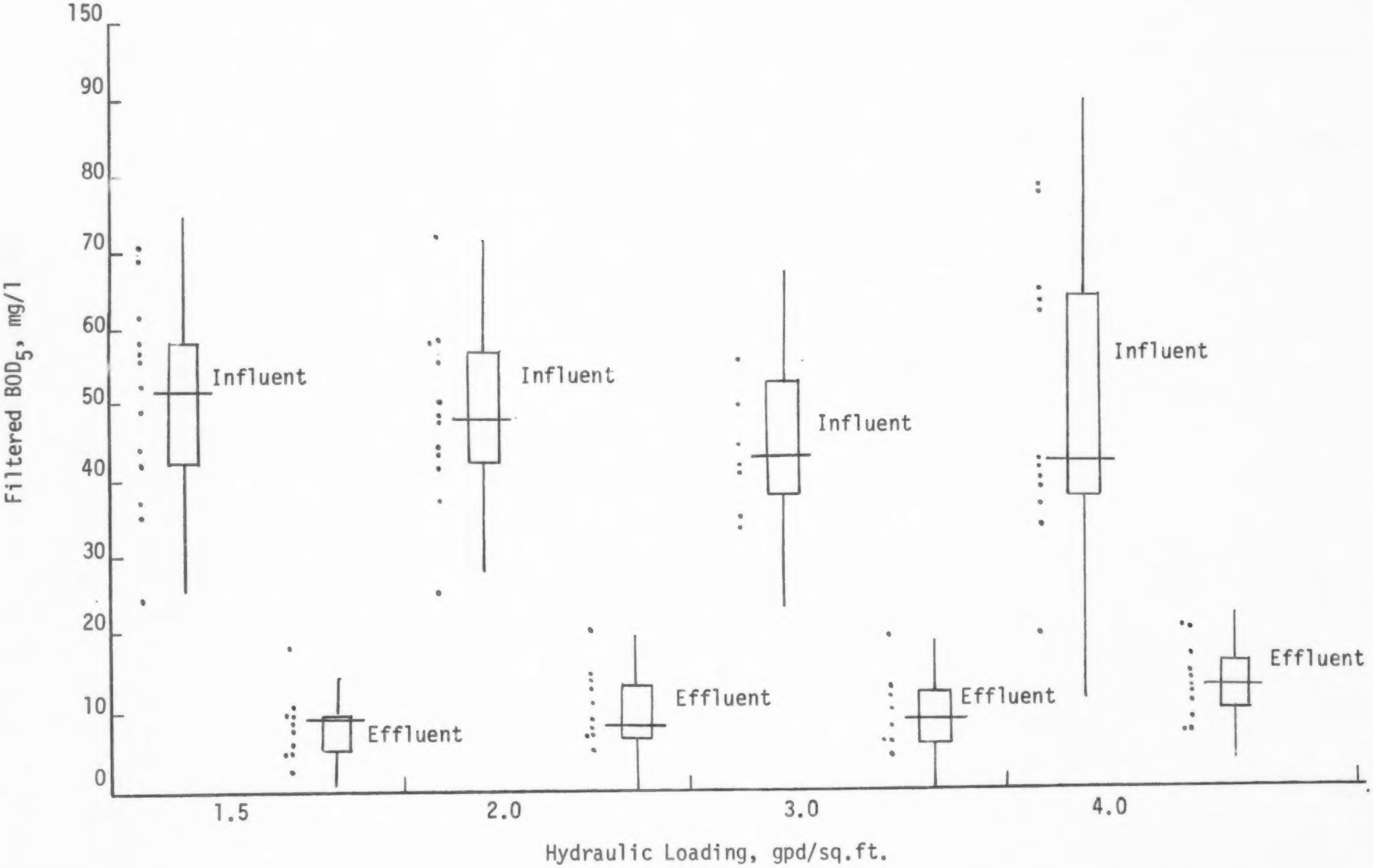

Figure 3. RBC Performance (for BOD_5 Removal) at Various Hydraulic Loading Rates. RBC influent concentrations for BOD are compared to RBC secondary effluent concentrations at each hydraulic loading rate, respectively.

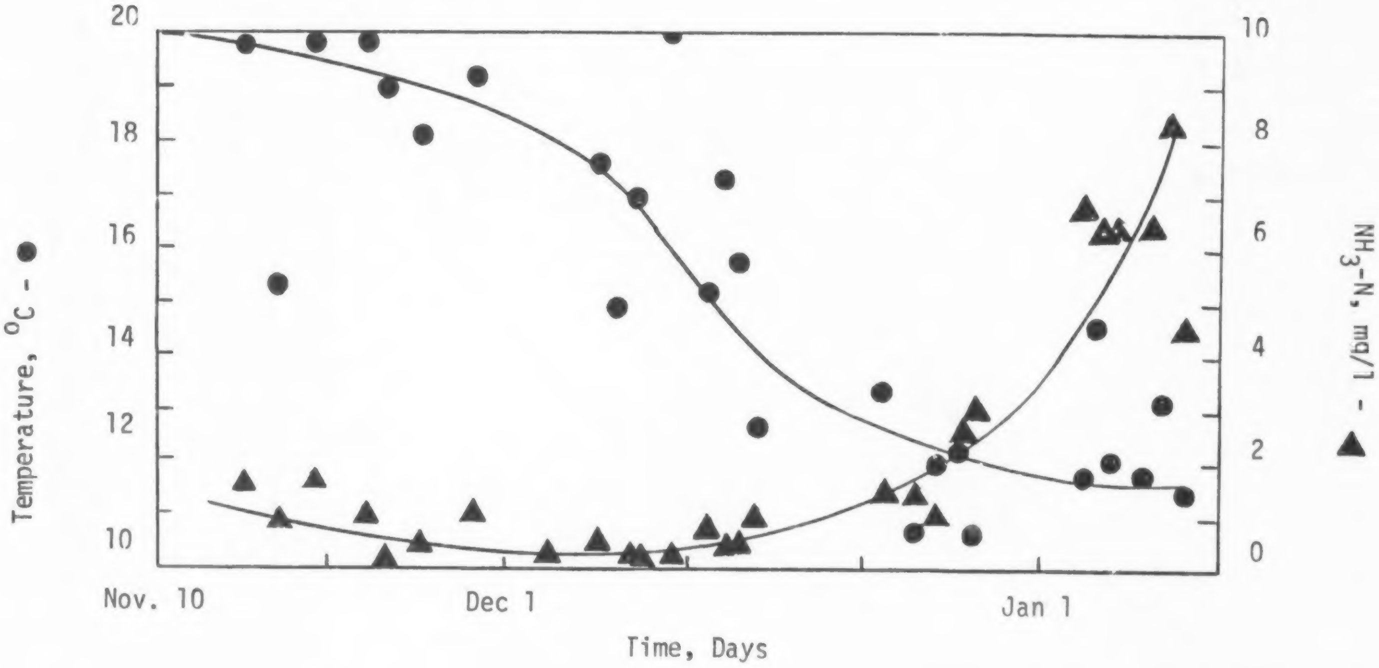

Figure 4. Effect of Temperature on RBC Effluent NH_3-N Concentrations at a Hydraulic Loading of 2.0 gpd/sq. ft.

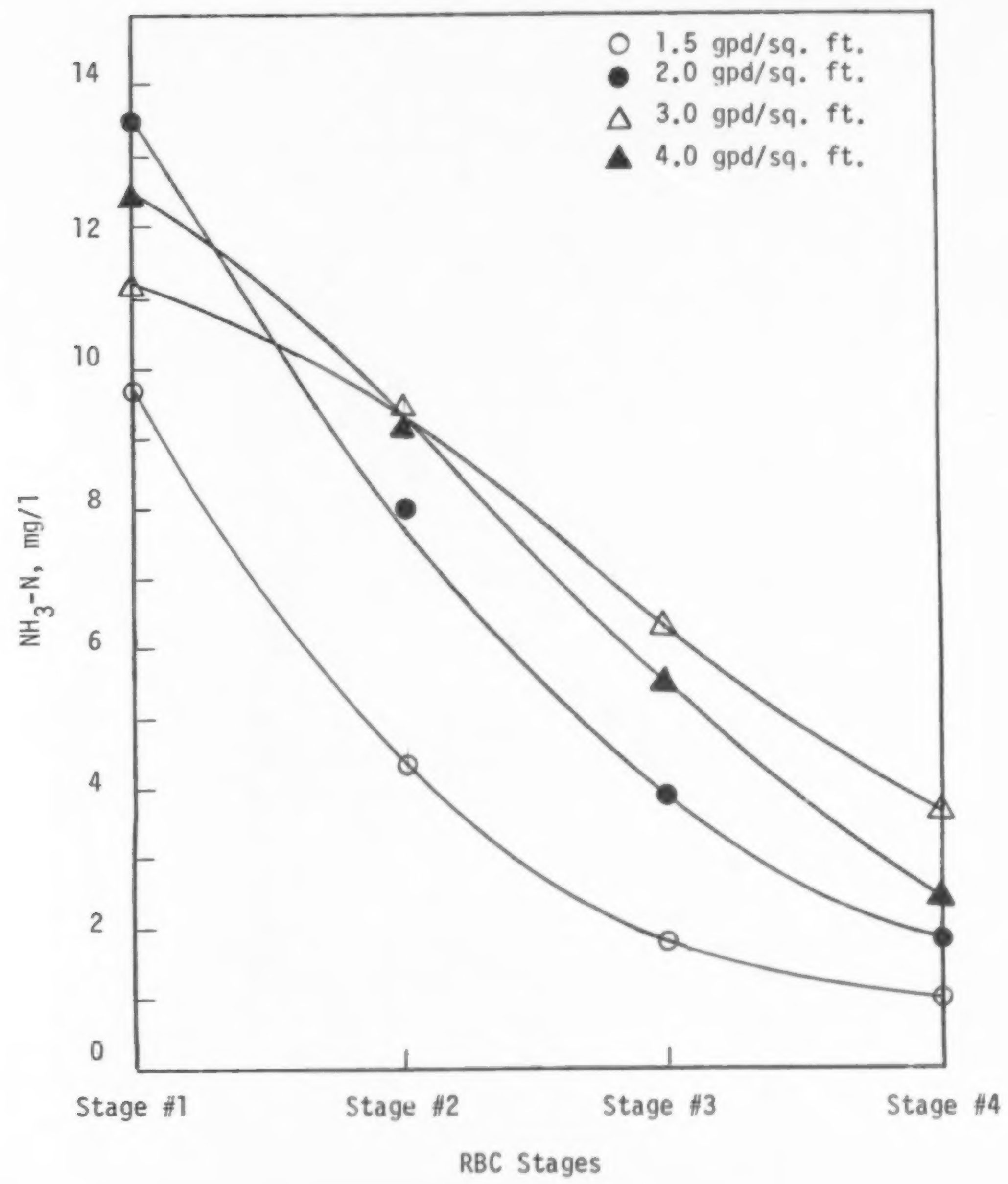

Figure 5. Progression of Nitrification within the RBC Process at Various Hydraulic Loadings.

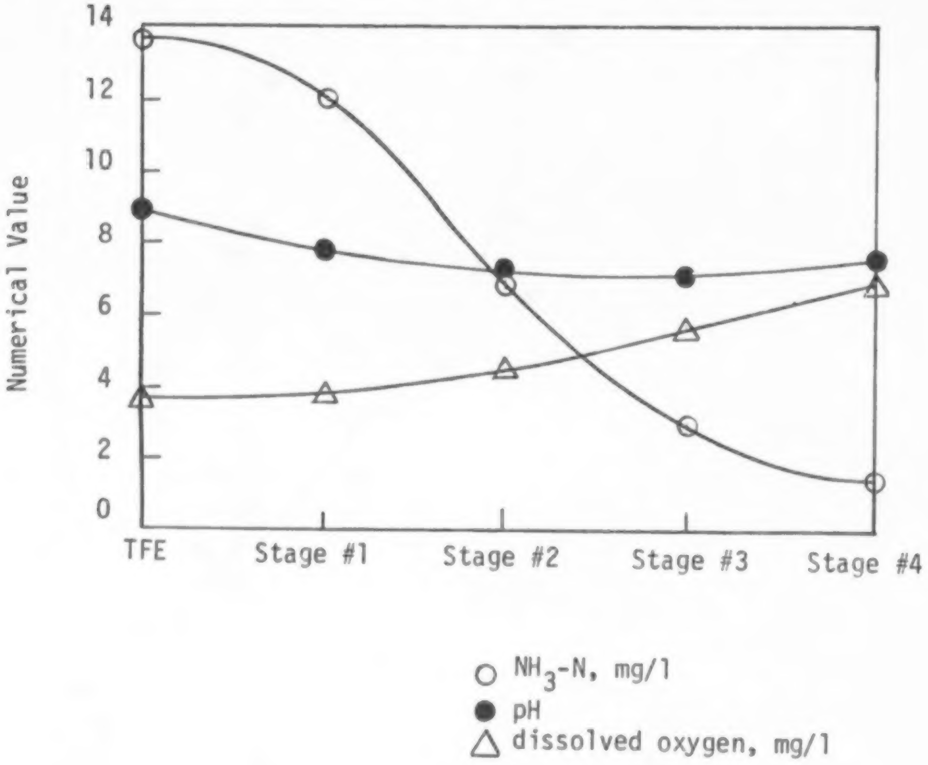

Figure 6. Progression of Nitrification within the RBC Process at 3.0 gpd/sq. ft. following a Trickling Filter without Intermediate Settling.

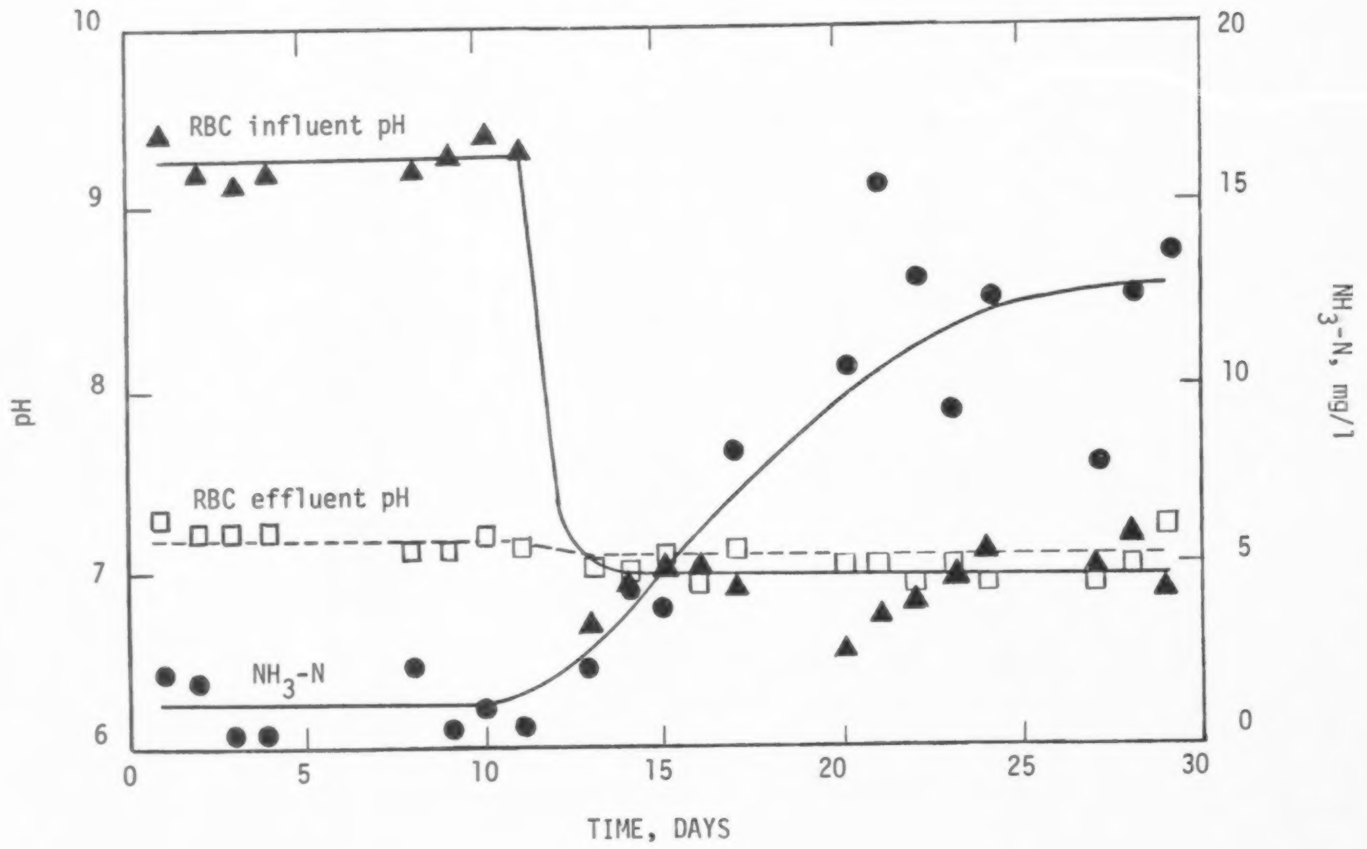

Figure 7. Effect of pH on Nitrification Across the RBC Process at 3.0 gpd/sq. ft.

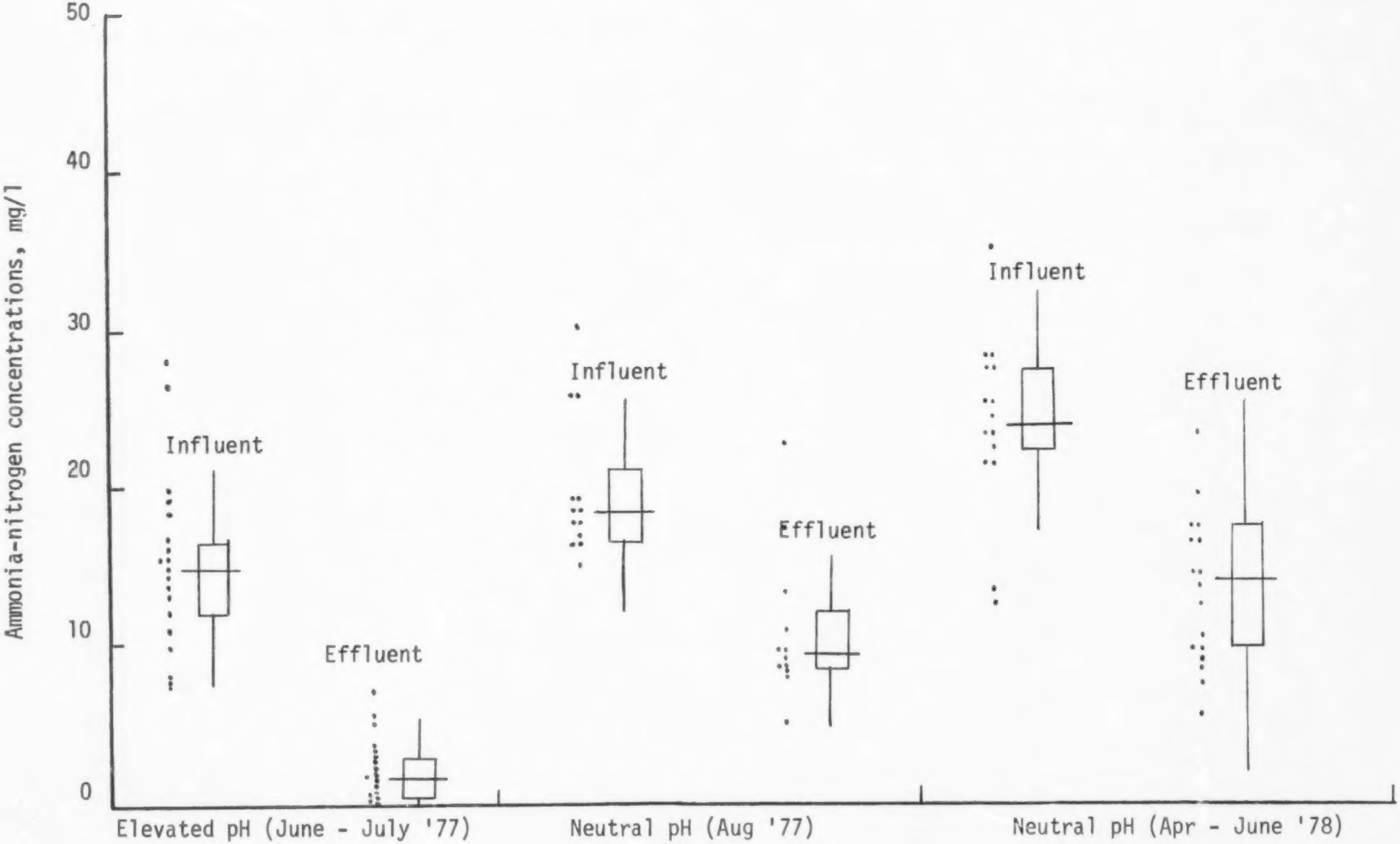

Figure 8. RBC Performance for Nitrification at 3.0 gpd/sq. ft. without Intermediate Settling. RBC influent and RBC secondary effluent ammonia-nitrogen concentrations are shown for elevated and neutral pH values.

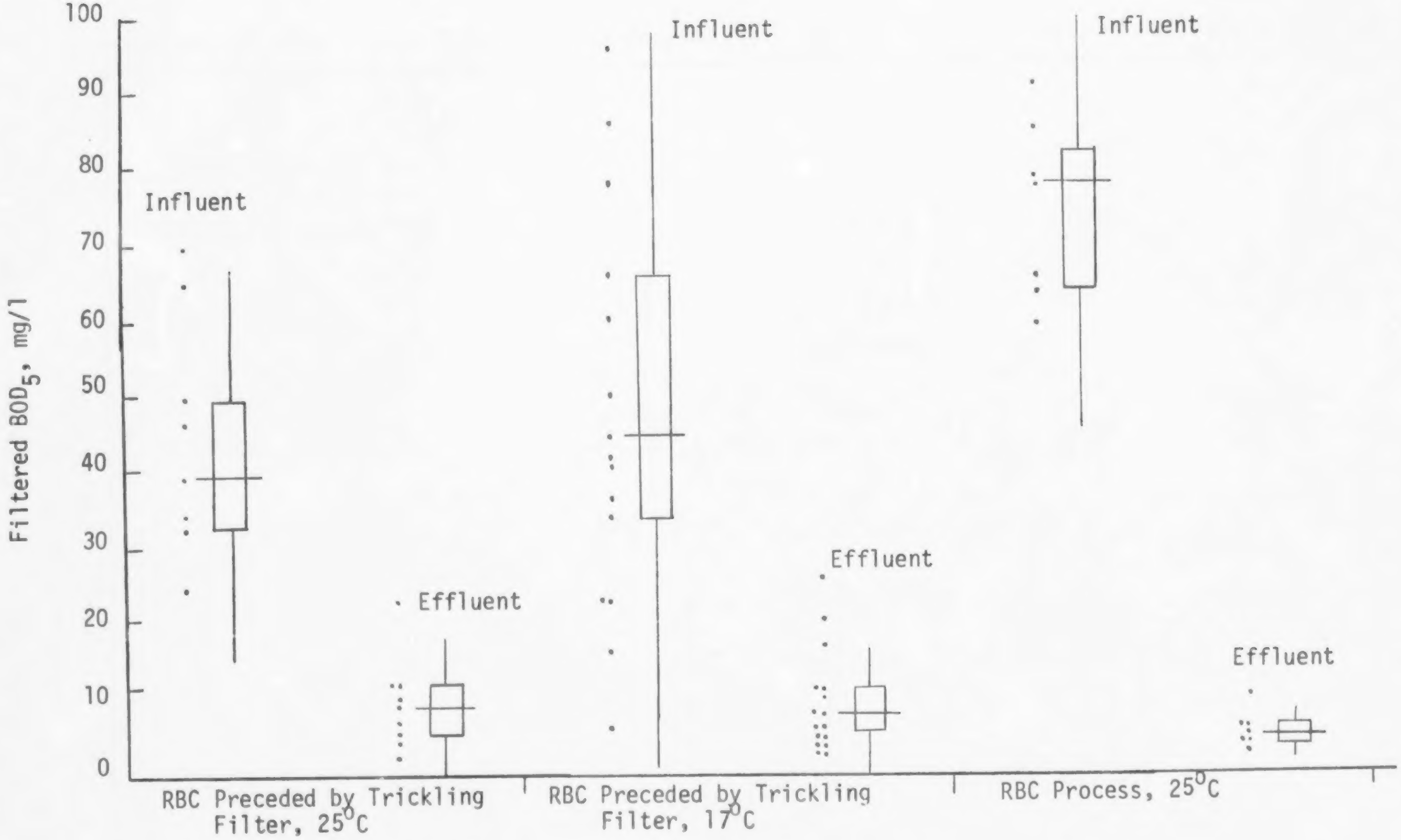

Figure 9. Secondary Treatment Performance of the RBC Process at 3.0 gpd/sq. ft.

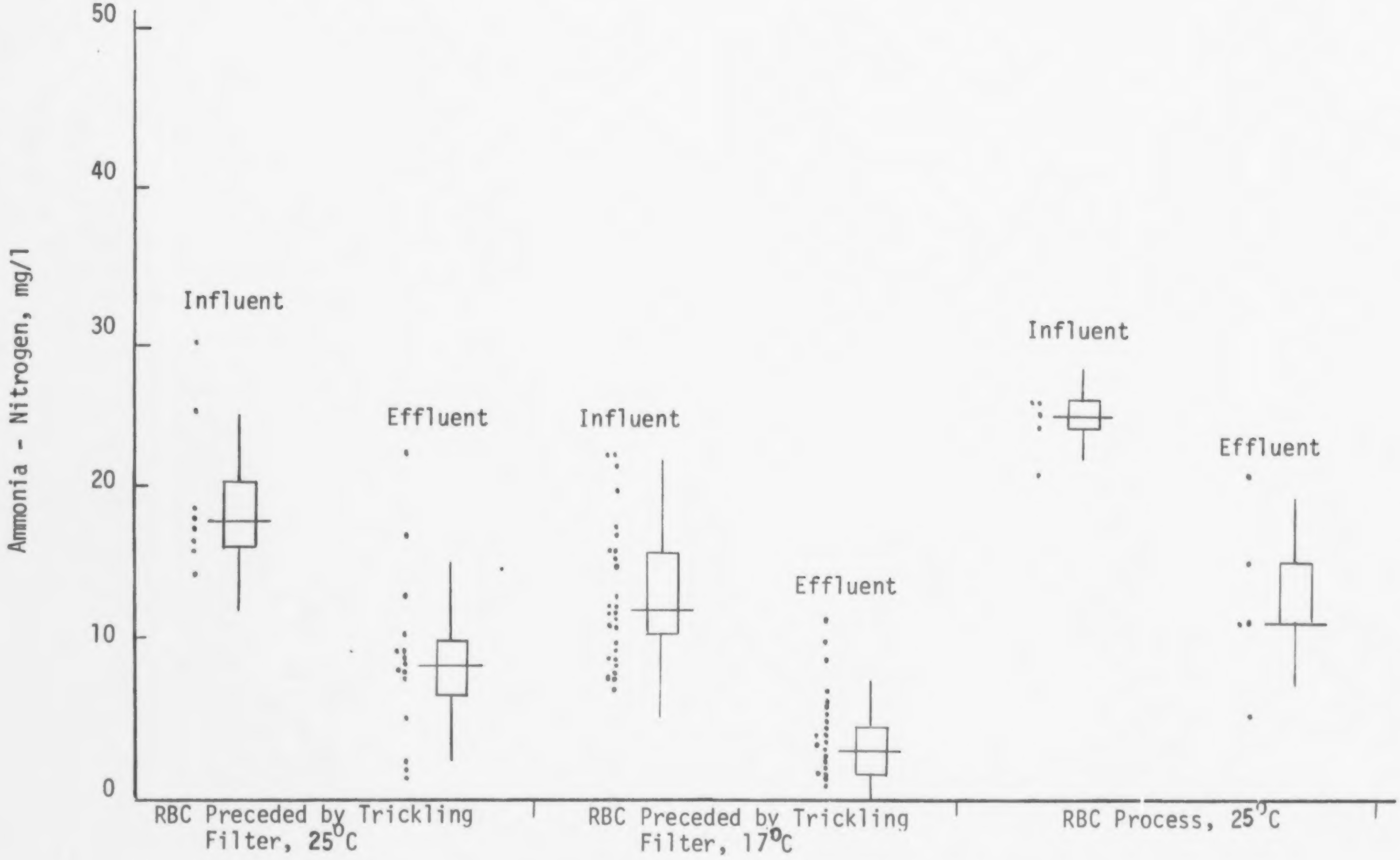

Figure 10. Nitrification Treatment Performance of the RBC Process at 3.0 gpd/sq. ft.

PART IX: DESIGN AND OPERATION

A DESIGN APPROACH FOR ROTATING BIOLOGICAL CONTACTORS TREATING INDUSTRIAL WASTEWATERS

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This paper presents a design model for rotating biological contactors treating industrial wastewaters based upon the multiple zero order organic removal concept. This same model has previously been applied to trickling filters and the activated sludge process. The basis for the model is that the organic removal rate in each stage is proportional to the concentration of organic remaining in that stage. Consider an industrial wastewater containing a mixture of organics of varying removal rates. The more readily degradable constituents will be removed first resulting in a higher overall removal rate in the early stages followed by a progressively decreasing removal rate in subsequent stages. This is defined by the relationship:

$$Q/A\ (S_o - S) = K\ S$$

in which Q = hydraulic loading gal/day (m^3/day)

A = filmed area ft^2 (m^2)

S_o = influent soluble BOD, mg/l

S = effluent soluble BOD, mg/l

K = proportionality constant

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A maximum rate will result from oxygen limiting conditions. Other models are compared for their application to various wastewaters.

A design example for an industrial wastewater is presented.

NOMENCLATURE

S_o = concentration in substrate of the influent ($mgBOD_5/l$)

S_i = concentration in substrate of the effluent of stage i ($mgBOD_5/l$)

n = number total of stage

Q = flow applied to the system of n stages (m^3/d)

A = wetted area of the discs in every stage (m^2)

ΣA = wetted area of the discs for the total system (n stages)

Q/A = hydraulic loading for every stage ($m^3/m^2 .d$)

k = kinetics constant (d^{-1})

K = proportionality constant ($m^3/m^2 d$)

K_f = saturation constant in the Monod relationship ($mgBOD_5/l$)

V = volume of the trough (m^3)

ρ = removal efficiency per stage (percent)

r = coefficient of correlation

c = maximum substrate removal per one stage ($g\ BOD_5/m^2 .d$)

DEVELOPMENT OF THE DESIGN MODEL

The design model is based upon the multiple zero order organic removal concept that has been applied to trickling filters and the activated sludge process treating domestic and industrial wastewaters (TISCHLER et al, 1969). The basis for the model is that the organic removal rate in the trough is proportional to the concentration of organic remaining in the trough. This is defined by the relationship:

$$\frac{dS}{dt} = kS \qquad (1)$$

Assuming complete mixing in the trough and the mass of fixed micro-organisms is proportional to the wetted area of the discs. A mass balance on the substrate for the trough, as shown in figure 1-a is given by:

$$\frac{dS}{dt} = V = Q\ (S_o - S_1) - K\ S_1\ A \qquad (2)$$

which at steady state becomes

$$Q/A\ (S_0 - S_1) = K\ S_1 \qquad (3)$$

Considerations of relationship (1) and (2) leads to the fact that in this model mass transport of oxygen and substrate, are not explicitly included and that organic removal by suspended micro-organism in the trough is negligible. This last assumption is reasonable if the detention time in the trough is short and the concentration of the biomass is low.

The relationship (2) and (3) is identical to the KORNEGAY model (1972), where the Monod relationship is used for the organic removal rate, if one assumes that S_1 is negligible with respect to the saturation constant K_f. In this model the constant K is the proportionality constant between the removal rate and the concentration remaining. Implicit in the definition is the fact that more readily degradable organics are removed first thereby resulting in a decreasing removal rate as the residual concentration decreases. Biofilm properties are incorporated in the coefficient. It is necessary to consider reactor hydraulics. It's known (HERBERT, 1964, KORNEGAY and ANDREW 1969) that a series of completely mixed reactors provide a valid approximation of a plug-flow system. This consideration is illustrated (See Fig. 2) by the response of a system of RBC's with two identical stages to a pulse input tracer (POPEL 1964). These considerations lead to the conclusion that relationship (3) is only applicable stage per stage, the general form of relation (3) for a system of n identical stages is:

$$\frac{Q}{A}\ (S_{i-1} - S_i) = K\ S_i \qquad (4)$$

At steady state the concentration of substrate S_i leaving stage i in terms of the influent concentration S_0 is given by:

$$S_i = \frac{S_0}{(1 + K\ A/Q)^i} \qquad (5)$$

It is important to note that in relationship (5) Q/A refers to each stage and not to the total area of the staged system.

The plug-flow mode provides a definite advantage for staged operation as indicated by consideration of relationship (5).

MODEL PREDICTIONS

For a series of identical RBC stages when flow is constant the relationship (5) shows that the removal efficiency per stage, ρ, is a constant. Adams et al (1975), Grau et al (1975) and Grady and Williams (1975) showed the effect of influent concentration for the activated sludge process treating highly variable strength wastewaters. The following development assumes a moderately constant strength wastewater.

This is defined by the relationship:

$$\log\ (S_i/S_0) = i\ \log\ (1 - \rho) \qquad (6)$$

A plot of the logarithm of S_i/S_0 versus the stage number, i, would give a linear plot as shown in Figure (3) with a slope equal to the removal efficiency, ρ. This graphical approach is very useful, as described later, to study the substrate concentration profile through a stage RBC system.

The multi-zero order assumption used in the model implies that the removal efficiency is independant of the feed concentration and that the primary design factor is the hydraulic loading rate per unit disc area. The effect of hydraulic loading rate, expressed in terms of specific contact area, is shown in Figure (4).

These curves are similar to the empirical relationship of POPEL (1964) for a fixed influent concentration and to the design charts of the STENGELIN Corporation widely used in Europe to design RBC plants.

The usual interpretation of this model is to plot the removal rate per unit of disc area $\frac{Q}{A}(S_{i-1} - S_i)$ versus S_i. This will give a linear plot as shown in Figure (5) with a slope equal to K.

A maximum removal rate can be expected when the biofilm growth is at a maximum rate under the experimental conditions. This maximum rate will result from oxygen or other limiting conditions.

When oxygen is limiting, the capacity of the unit can be increased by increasing the rotational speed WELCH (1968). This limitation is generally observed at high loadings in the first stages in a staged RBC system. For cost optimization, however, one must consider that the power consumption increases exponentially with rotational speed.

MODEL VERIFICATION

The model has been tested with data available in the literature by ECKENFELDER (1977) and recently by EDELINE and VANDEVENNE (1979).

The multiple zero order assumption, relationship (6) has been verified in several cases, TORPEY (1971), BIRKS (1971), LABELLA (1972) and KINCANNON (1974) in Figures 6 to 9.

We observed in some cases a discontinuity in the plot for the first stage, where the removal rate is higher than that predicted from relationship (4). In this case as indicated in Figures 8 and 9, the first stage must be considered separately. A second discontinuity is also observed in some cases for the last stages of a RBC system where the removal rate decreases. This expresses a quantitative and/or qualitative modification of the fixed biomass, probably due to a variation in the nature of the substrate. This leads to a modification of the proportionality constant, K.

These considerations result from the staged system. The more readily degradable constituents are removed first resulting in a higher removal rate in the early stages followed by a progressively decreasing removal rate in subsequent stages. Typical results are presented in Figures 10 and 11.

It has been observed (EDELINE and VANDEVENNE, 1979) that when one only uses cumulative data, $S_0 - S_n$, these data can be interpreted by the model in some cases. In cases this is an empirical relationship that does not justify an extrapolation for a staged system with a variable number of stages. The value of K in this case is higher than the true value because it includes the plug-flow kinetics effect and eventually a higher removal in the first stage. The true value of K can be obtained by linearisation of relationship (5) as indicated in Figure 12. This fact has been also observed by the authors with the Kornegay model.

The results are summarized in Table 1 for domestic wastewater and in Table 2 for industrial wastewaters A better coefficient of correlation could be obtained in some cases if one had not forced the line to pass through zero. The residual line give a BOD_5 removal limited to an effluent BOD_5 from 3 to 10, which is reasonable when one considers the dynamics of the biological process.

COMPARISON WITH OTHERS MODELS

EDELINE and VANDEVENNE (1979) have compared this model with the Kornegay mode. In may cases the Kornegay model gives a better coefficient of correlation (plotting $\frac{A}{QAS}$ versus 1/S). However, the proposed model offers the simplicity of having only one constant.

When the flow is constant, the design model presented in this paper gives identical predictions as the model of HANSFORD et at (1978), that consider the biological reaction and mass transfer to the biofilm, since the authors assume fist order kinetics that has the same formulation as relationship (1). When the hydraulic loading varies a comparison of the results published by the authors from the ANTONIE et al (1971) data as presented in Figure 13, the predictions appear similar (the coefficient of correlation is not given by the authors). This may imply that mass transfer processes are not influenced by a modification of flow applied to the RBC. This fact can be related to the results of KORNEGAY (1972) that showed that flow rate does not have as significant effect on the constant of saturation.

DESIGN EXAMPLE

The following data was obtained on a RBC for a bleached pulp mill effluent. The test unit had 4 stages.

Flow to the unit		*BOD_5 (mgO_2/l)*				
		Influent	*Stage 1*	*Stage 2*	*Stage 3*	*Stage 4*
g pd/ft^2	*m^3/m^2 d*					
2	0,082	76	26	10	7	6
4	0,163	72	41	24	16	12
6	0,245	81	42	25	15	10
8	0,326	78	52	38	25	18
10	0,408	83	60	47	36	30

Temperature : 18 °C

TABLE 1

DOMESTIC WASTEWATERS

Reference	Temperature (°C)	K m^3/m^2 d	C g $\Delta BOD_5/m^2$ d	r
Antonie and Koehler	12,5-20	0,204	-	0,9059
Ban Huat Tan	(1)	0,633	35	0,8747
Clark et al	(1)	0,301	-	0,7420
Marki	(1)	0,410	-	0,6640
Põpel	(1)	0,428	32	0,8012
Stengelin	(1)	0,679	105	-
Torpey	16 - 26	0,701	-	0,9404

(1) In these studies temperature follow seasonal variation

Note :-with flow in gpd/ft^2, multiply the value of K (m^3/m^2 d) by 24,51.

-°F = 32° + 5/9 °C

TABLE 2

INDUSTRIAL WASTEWATERS

References	Wastewater	Temperature (°C)	K m^3/m^2 d	C g BOD_5/m^2 d	r
Antonie et Welch	Dairy	12-20	0,082*	-	0,8909
Birks et al	Cheese	7-20	0,039	24	0,9543
Bintanja	Synthetic substrate (caseine)	20	0,055*	36*	0,7998
Cebedeau	Malte-house	20	0,037	-	0,4634
Chen et al	Paper-mill	30-35	0,177	-	0,5282
Chittenden et al	Meat (pretreated by anaerobic lagoon)	(20)	0,343	-	0,8182
Cochrane et al	Potato	15-18	0,082	40	0,8209
Eckenfelder	Bleached pulp mill	18	0,556	-	0,9525
Kincannon et al	Meat	-	0,026	60	0,9050
Labella et al	Winery	-	0,126 0,150(6stages)	170	0,9485 0,9150
Thomas et al	Distillery	-	0,435**	-	0,9647
Willard et al	Hardboard	14-19	0,056	-	0,9779

* Data in DCO

** Relative to the total RBC unit.

Note :-with flow in gpd/ft^2, multiply the value of K (m^3/m^2 d) by 24,51
- °F = 32° + 5/9 °C

The proportionality constant is calculated by plotting Q/A $(S_{i-1} - S_i)$ versus S_i.

The hydraulic load per stage, Q/A, is given by multiplying the hydraulic load relative to the unit by 4.

From Figure 14, K = 0.556 m^3/m^2d (13.6 gpd/sq ft)

Design an RBC plant for 80 percent BOD removal assuming an influent BOD_5 of 85 mg/l and a required effluent BOD of 17 mg/l.

The corresponding Q/A to meet S_e = 17 mg/l can be calculated by the relationship:

$$\frac{S_n}{S_o} = \frac{1}{(1 + K\ A/Q)^n}$$

with S_n = 17 mg/l

n = number of stages of the RBC unit

The results are plottted in Figure 5.

The total area of discs is calculated based on the hydraulic load for the total system, $Q/_{\Sigma A}$:

$$\frac{Q}{\Sigma A} = \frac{Q}{A} \cdot \frac{1}{(n)}$$

where n is the number of stages in the RBC unit. These results are plotted in Figure 16.

To meet S_e = 17 mg/l in a three stage RBC system, the hydraulic load for the system is 0.26 $m^3/m^2.d$ (6.37 gpd/sq ft).

If the influent flow is 2,000 m^3/d (0.528 mgd) the total surface area is:

$$\frac{2{,}000\ m^3/d}{0.26\ m^3/m^2.d} = 7{,}692\ m^2\ (\approx 82.800\ \text{sq ft})$$

The final number of stages can be decided by a cost optimization.

When the experimental data shows a maximum removal capacity, the organic loading for each stage should be checked. If this loading is higher than the maximum, the hydraulic loading should be decreased.

Also when one observes a high removal rate in the first stage, this stage should be considered separately and the other stages calculated as indicated in the example.

SUMMARY AND CONCLUSIONS

A model based on a multi-zero order kinetics, is developed and the method for evaluating the constant described. Validations and limitations are described on the basis of published data. Constants are given for domestic and some industrial wastewaters.

The simple relationship reasonably fits experimental data and can be used to design rotating biological contactors. In some cases, special emphasis on the first stage removal is necessary. A maximum removal that may be due to oxygen limitation must be considered also.

At this time, several models may describe the removal relationship in a staged RBC unit. For this reason, the authors recognize that in some cases empirical relationships or design charts may be better for the design. It is essential to record during a pilot plant study the substrate concentration profile in a staged RBC unit.

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EMPIRICAL OXYGEN TRANSFER RELATION IN THE RBC PROCESS

by

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INTRODUCTION

Maintenance of bulk solution dissolved oxygen concentrations in the Rotating Biological Contactor process (RBC) has been found to be important in retaining process efficiency in treating simulated dairy wastes (1), synthetic sewage (2), petroleum wastes (3), and in nitrification (4). Two hypotheses can be made to explain the observed phenomenon. The first would suggest a loss of efficiency of metabolism to low oxygen concentrations. The second would suggest a population change occurs on the disc such that organisms which can withstand low oxygen concentrations begin to predominate resulting in a net loss of process efficiency. Dissolved oxygen concentration limits have been observed to be between 1.0 to 2.1 mg/l. The lack of understanding of oxygen transfer in pilot and full scale RBC plants has apparently resulted in

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cases where the full scale plant was unable to meet the predicted removal efficiency (5).

Two procedures have been suggested to scale-up oxygen transfer effects in pilot and full scale RBC systems. Oxygen transfer to the bulk solution by rotating discs has been presented as a function of disc surface area turnover (2, 5, 6). This concept has been used to argue for pilot plant scale-up based on disc rotational speed (rpm; revolutions per minute). However, other literature suggests pilot plant scale-up based on disc tip speed as the design parameter (7).

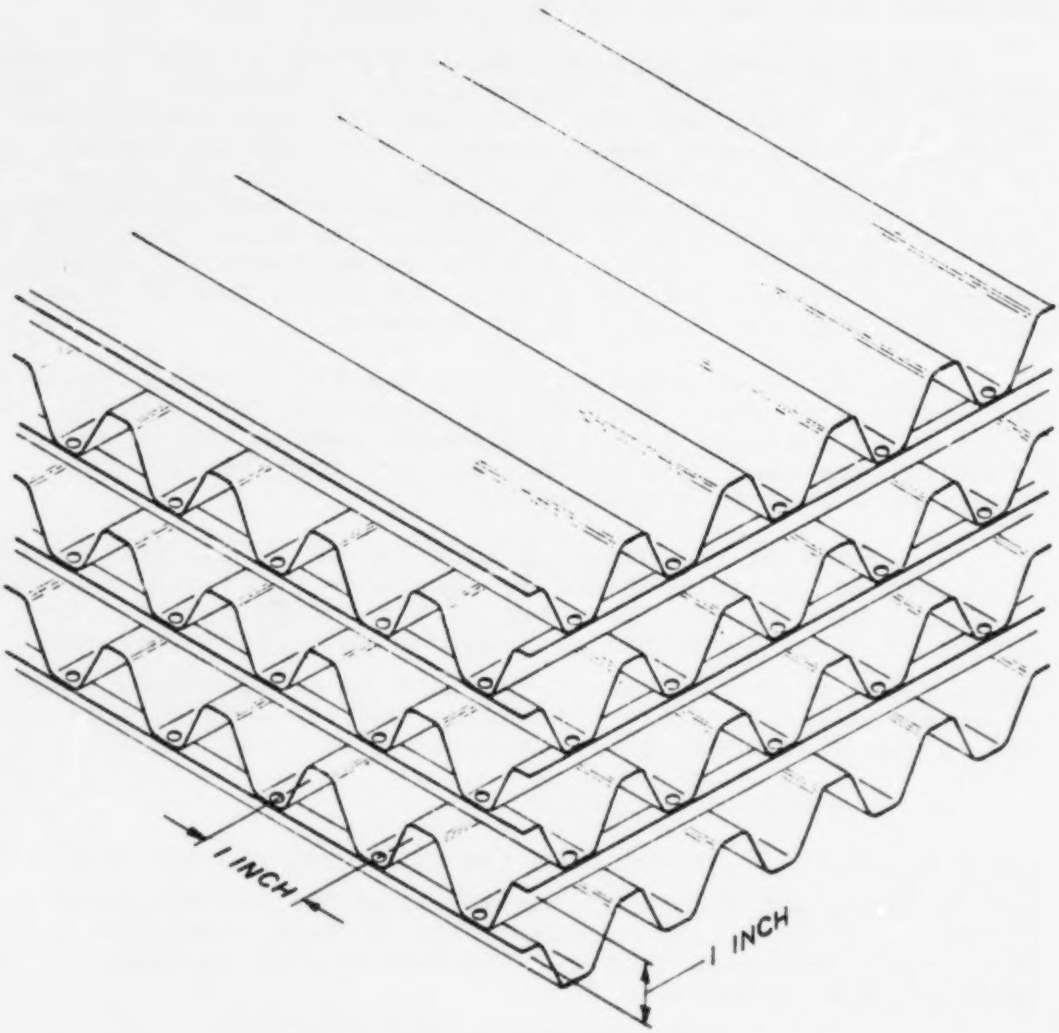
Oxygen transfer to the bulk solution is governed by water adhering to the discs, as well as mixing in the tank. Neither of the two models for scale-up, i.e., tip speed and rpm, fully consider the role of tank mixing in transfering oxygen. In order to understand the role of tank mixing in the transfer of oxygen, it is necessary to model the effects of tank volume and geometry on oxygen transfer. Zeevalkink et al. (8) used an empirical relation based on disc submergence depth to correlate a series of oxygen transfer data taken with flat, polystyrene discs.

Data presented in this report show that neither tip-speed nor rpm scale-up adequately defines oxygen transfer relations between pilot RBC systems tested with clean tap water and clean discs. An empirical relation between oxygen mass transfer and Reynolds numbers calculated from tank geometries is developed. A scale-up factor relating rotational effects in pilot and full-scale systems is proposed.

RBC PILOT PLANTS

Oxygen transfer was evaluated in three RBC pilot plants; Bensenville, Illinois; Princeton, New Jersey; and Santa Clara, California. All RBC discs were made with the FMC Corporation, Environmental Equipment Division, corregated media design shown in Figure 1. A total of eight test cases at the three pilot facilities were studied for the development of the oxygen transfer model. A test case is defined as a variation in the number of discs, the water depth in a tank, or an alteration in the distance between the bottom of the tank and the bottom edge of the discs.

The Bensenville RBC pilot plant, located at the South Wastewater Treatment Plant, Bensenville, Illinois, is a full size, twenty-five foot shaft, with media divided into four stages (66 sheets of media per stage). The full shaft carries 84,000 square feet of media. The discs are 11.69 ft. in



1 INCH
1 INCH

MEDIA DETAIL
HIGH DENSITY POLYETHYLENE
FIGURE 1

diameter. Oxygen transfer tests were made in the first and third stage of this pilot plant. Data presented are the average of the two results and, therefore, represent a single stage of the four stage system. Two test cases were studied at this pilot plant. In test case B the water depth was maintained at the normal operating depth. In test case BL the water level was lowered by six inches.

Four test cases were studied at the FMC Central Engineering Laboratory facility in Santa Clara, California. In the first three test cases, C10, C22, and C30, ten, twenty-four and thirty discs were used, respectively. In the fourth test case, C30H, the shaft containing 30 discs was raised by five inches to increase the distance between the bottom of the tank and the edge of the media. The water level was also raised by five inches in order to maintain the same percent submergence of the discs. The media used in this pilot facility was 11.83 feet in diameter.

Two test cases were studied at the FMC Chemical Research and Development pilot facility at Princeton, New Jersey. This pilot plant utilized 3.95 ft. diameter media. Eight sheets of media were used on the shaft in test case P8, and ten sheets were used in test case P10.

A summary of the physical data for all eight test cases is presented in Tables 1 and 2. In Table 1 the disc diameter is represented by (2 x r) where r is the disc radius. The number of discs is n, the rank volume is V_t, and the tank water depth at the center of the tank is H. The tank width is given as w. In table 2 the disc section length is L. The active disc area, a, is calculated in Equation 1:

1) $a = mn\pi\ (r^2 - r_w^2)$

where r_w is the distance from the center of the RBC shaft to the water and m is a factor relating the surface area of one sheet of plastic media to the surface area of a circle with the same radius. The active disc volume, V_d, is expressed in Equation 2:

2) $V_d = fL\pi\ (r^2 - r_w^2)$

where f is the void volume of the media. The active volume and the active area are the volume and area of media which become submerged during one revolution of the shaft.

TABLE 1: PHYSICAL DATA FOR RBC PILOT PLANTS

PILOT PLANT	(2 x r) DISC DIAMETER (ft)	(n) NUMBER OF DISCS	DIAMETER SUBMERGENCE (PERCENT)	(V_t) TANK VOLUME (ft^3)	(H) TANK WATER DEPTH (ft)	(w) TANK WIDTH (ft)
B	11.69	66	37.3	326.8	5.25	6.72
BL	11.69	66	33.0	282.9	4.75	6.72
C10	11.83	10	37.5	137.6	4.67	3.33
C22	11.83	22	37.5	135.3	4.67	3.33
C30	11.83	30	37.5	133.7	4.67	3.33
C30H	11.83	30	37.5	156.7	5.08	3.33
P8	3.95	8	38.4	5.25	1.75	1.17
P10	3.95	10	38.4	5.20	1.75	1.17

TABLE 2: PHYSICAL DATA FOR RBC PILOT PLANTS: CONTINUED

PILOT PLANT	(L) DISC SECTION WIDTH (ft)	(a) DISC AREA (ft^2)	(V_d) ACTIVE DISC VOLUME (ft^3)	(E) DISTANCE DISC EDGE TO TANK BOTTOM (ft)
B	5.50	20070	517.8	0.896
BL	5.50	19007	490.3	0.896
C10	0.83	3094	79.8	0.250
C22	1.83	6807	175.6	0.250
C30	2.50	9282	239.5	0.250
C30H	2.50	9282	239.5	0.670
P8	0.67	284	7.33	0.167
P10	0.83	355	9.16	0.167

The tanks at Bensenville and Santa Clara were semicircular in design. The tank at Princeton was an angular design, such that the distance between the media and the tank wall was fairly constant. The Princeton tank, therefore, approached a semicircular design. The distances between the tank and media for all test cases is approximated by the distance, E, in Table 2.

METHODS AND PROCEDURES

Oxygen transfer was evaluated using clean RBC discs in clean tap water with the non-steady state aeration test. The dissolved oxygen was first removed by the addition of approximately 120 mg/l sodium sulfite. Cobalt chloride (0.5 mg Co/l) was used as a catalyst. As the discs rotated, the oxygen content of the test water slowly increased. Oxygen concentrations were determined with YSI dissolved oxygen probes and meters (Yellow Springs Instrument Co., Yellow Springs, OH) in tests conducted at the Princeton and Santa Clara pilot plants. The modified (sodium azide) Winkler dissolved oxygen test (9) was used to determine oxygen concentrations at the Bensenville pilot plant. The oxygen saturation value was determined, in most cases, by allowing the test to proceed until a constant dissolved oxygen concentration was achieved. In several cases the saturation concentration was determined from tabulated values in Standard Methods (9). The oxygen transfer coefficient K_La, for each test was calculated as the negative slope of a plot of the natural logarithm of the oxygen deficit versus time. The oxygen deficit is the difference between the observed oxygen saturation concentration and the oxygen concentration at the sample time. Data representing between 10 and 90 percent of saturation were used to calculate K_La. All tests were performed with water temperatures near 20°C. Corrections of K_La to K_La_{20}, the oxygen transfer coefficient at 20°C, were made using the mathematical procedure discussed in Standard Methods (9).

RESULTS AND ANALYSIS

Oxygen transfer coefficients, K_La_{20} (min^{-1}), at several rpm values, were determined for each of the eight test cases. Oxygen transfer rates, OTR (gO_2/min) were calculated for each test using Equation 3 and the proper unit conversions:

$$3) \quad OTR = K_La_{20}V_tC_s$$

where C_s is the saturation value of oxygen at 20°C (9.2 mg O_2/ℓ). The results for K_La_{20} and OTR are reported in Table 3.

TABLE 3: OXYGEN TRANSFER RESULTS

PILOT PLANT	rpm	K_La_{20} (min^{-1})	OTR OXYGEN TRANSFER RATE (gO_2/min)
B	0.8	0.0170	1.44
	1.2	0.0265	2.26
	1.6	0.0450	3.83
	2.4	0.1150	9.77
BL	1.2	0.0315	2.32
	1.6	0.0521	3.84
	2.4	0.1314	9.67
C10	1.52	0.0135	0.484
	1.71	0.0155	0.555
	2.26	0.0245	0.878
C22	1.52	0.0323	1.138
	1.71	0.0399	1.406
	1.88	0.0507	1.787
C30	1.52	0.0361	1.26
	1.71	0.0471	1.64
	1.88	0.0622	2.17
C30H	1.52	0.0339	1.38
	1.71	0.0444	1.81
	1.88	0.0594	2.42
P8	1.2	0.011	0.015
	3.0	0.048	0.066
	5.0	0.101	0.138
	7.5	0.272	0.374
P10	1.2	0.019	0.026
	3.0	0.059	0.080
	5.0	0.133	0.180
	7.5	0.314	0.425

Several attempts were made to linearize the oxygen transfer results from the eight test cases. None of the attempts was successful until a method of estimating fluid mixing was developed. This method is based on a Reynolds number calculated from tank geometry. The Reynolds number is developed by considering a reference cross-section of the RBC tank and media. The reference cross-section is a plane observed parallel through the shaft and perpendicular to the ground (Figure 2). The Reynolds number, Re, is calculated using Equation 4:

$$4)\quad Re = \bar{v}R/\upsilon$$

where the net water velocity through the reference cross-section is $\bar{v}$ (ft/min), the hydraulic radius of the cross-section is R (ft), and υ is the kinematic viscosity.

Several assumptions are required in order to calculate the Reynolds number. The first assumption is that the media acts as a pump to move water from one side of the tank to the other across the reference cross-section. An estimate of the pumping capacity of the media, Q(ft^3/min), is given by Equation 5.

$$5)\quad Q = rpm \times V_d$$

A second assumption is that all of the water pumped across the reference cross-section must flow back to the other side of the tank. The media is viewed as an impassable barrier to water flowing in the direction opposite to the disc rotation. All the flow must pass around the media through the area, A (ft^2), between the tank wall and the media. This area is calculated using Equation 6.

$$6)\quad A = (E \times L) + H \times (W - L)$$

The estimated net water velocity, $\bar{v}$ (ft/min), is given by Equation 7.

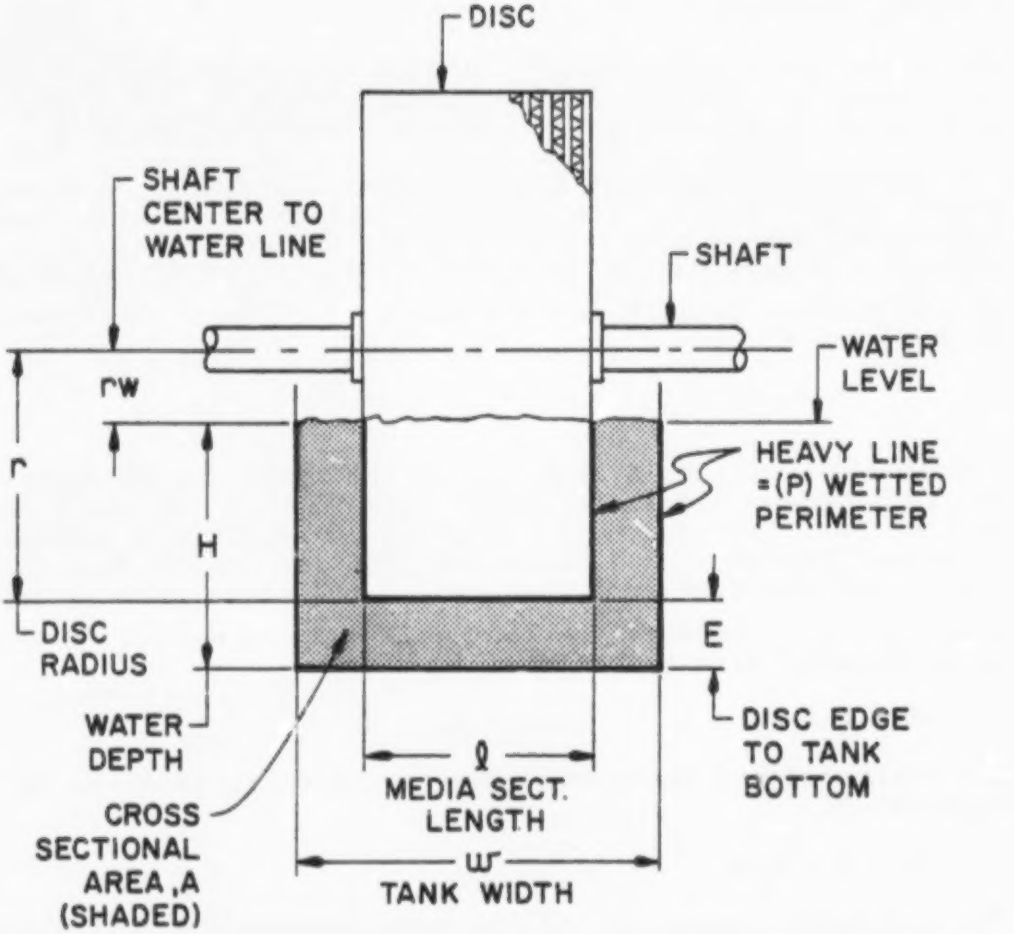
$$7)\quad \bar{v} = Q/A$$

The hydraulic radius, R, of the area between the tank wall and the media is found using Equation 8:

$$8)\quad R = A/p$$

where p (ft) is the wetted perimeter of the reference cross-section calculated using Equation 9.

$$9)\quad p = 4H - 2E + w + L$$


DISC
SHAFT CENTER TO WATER LINE
SHAFT
WATER LEVEL
rw
r
H
HEAVY LINE =(P) WETTED PERIMETER
E
DISC RADIUS
WATER DEPTH
DISC EDGE TO TANK BOTTOM
l
MEDIA SECT. LENGTH
CROSS SECTIONAL AREA, A (SHADED)
w
TANK WIDTH

CROSS-SECTION REFERENCE VIEW

FIGURE 2

Table 4 presents the data required to calculate Reynolds numbers for each test case.

Calculation of a dimensionless parameter to describe the oxygen transfer data was fashioned after Wu, et al., (10). These authors used a dimensionless parameter of the form given in Equation 10 to describe oxygen transfer by mechanical mixers.

$$10)\quad N_{20} = OTR/\mu d$$

The dimensionless oxygen transfer number, N_{20}, is calculated for data standardized to 20°C. The dynamic viscosity is μ, and d is a characteristic distance in the reactor design.

A characteristic distance, d, for correlating RBC data was found to be the volume of the tank divided by the active disc area (Equation 11).

$$11)\quad d = V_t/a$$

Equation 10 can be modified by replacing expressions for OTR and d. This is shown in Equation 12.

$$12)\quad N_{20} = \frac{K_L a_{20}\, C_s\, a}{\mu}$$

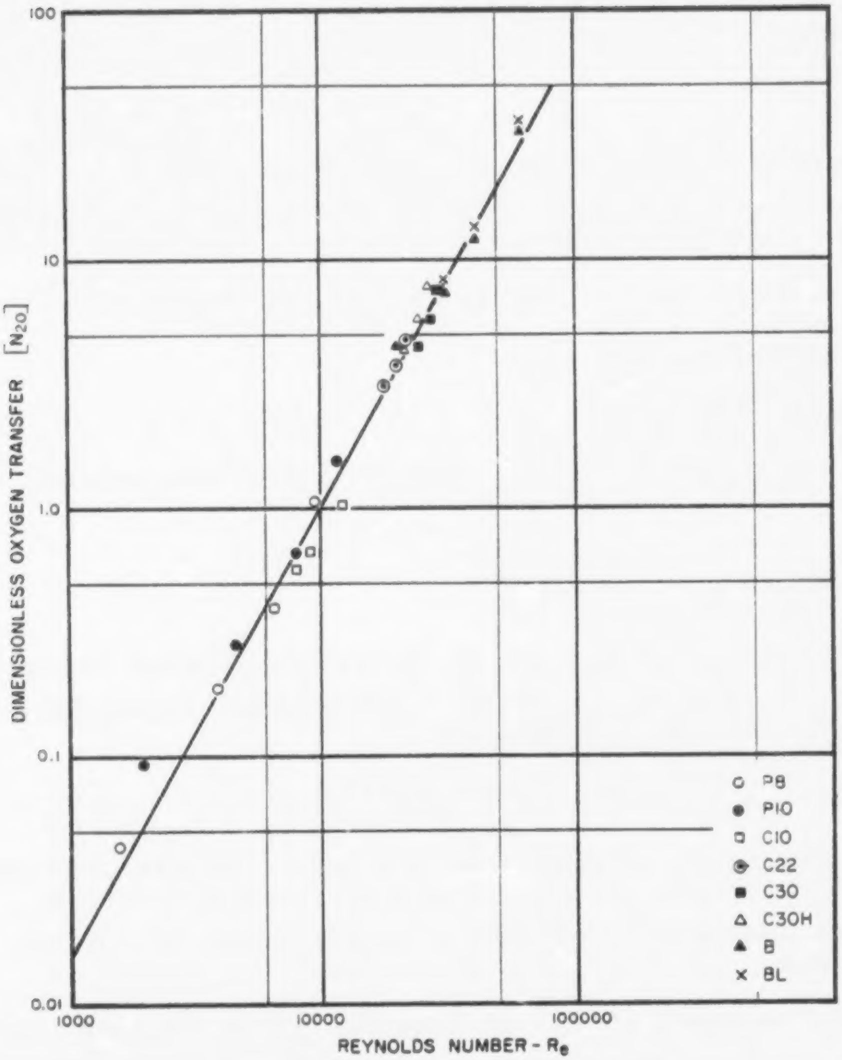
A summary of N_{20} and Re values calculated for all the test data is given in Table 5 and plotted in Figure 3. The plotted line has the equation:

$$13)\quad N_{20} = 5.7 \times 10^{-8}\ Re^{1.815}$$

The correlation coefficient is 0.994. Further development of dimensionless parameters was not pursued since N_{20} and Re were adequate to produce a linear relation for the available data.

Replacement of N_{20} in Equation 13 by the original parameters outlined in Equations 1 and 12 and replacement of Re in Equation 13 by the original parameters outlined in Equations 2, 4, 6, 7, and 9, results in an equation expressed in readily measured terms (Equation 14).

$$14)\quad \frac{K_L a_{20} C_s \pi\, mn(r^2 - r_w^{\ 2})}{\mu} = 5.7 \times 10^{-8}\ \left(\frac{\pi f L}{\upsilon p}\ rpm\ (r^2 - r_w^{\ 2})\right)^{1.815}$$



DIMENSIONLESS OXYGEN TRANSFER
VERSUS REYNOLDS NUMBERS

FIGURE 3

TABLE 4: REQUIRED INFORMATION FOR REYNOLDS NUMBER CALCULATIONS

PILOT PLANT	($\bar{v}$) WATER VELOCITY (ft/min)	(R) HYDRAULIC RADIUS (ft)	(A) CROSS SECTIONAL AREA (ft^2)	(p) WETTED PERIMETER (ft)
B	45.65 x rpm	0.362	11.34	31.4
BL	45.73 x rpm	0.365	10.73	29.4
C10	6.69 x rpm	0.531	11.88	22.3
C22	23.07 x rpm	0.326	7.46	23.3
C30	50.74 x rpm	0.195	4.51	24.0
C30H	39.20 x rpm	0.236	5.90	25.7
P8	7.51 x rpm	0.116	0.98	8.4
P10	12.57 x rpm	0.084	0.73	8.7

TABLE 5: DIMENSIONLESS OXYGEN TRANSFER AND REYNOLDS NUMBERS

PILOT PLANT	rpm	N_{20}	Re
B	0.8	4.63	20200
	1.2	7.22	30300
	1.6	12.26	40400
	2.4	31.33	60600
BL	1.2	8.13	30530
	1.6	13.44	40700
	2.4	33.90	61060
C10	1.52	0.57	8320
	1.71	0.65	9360
	2.26	1.03	12370
C22	1.52	2.98	17490
	1.71	3.69	19680
	1.88	4.68	21640
C30	1.52	4.55	23190
	1.71	5.93	26090
	1.88	7.84	28680
C30H	1.52	4.27	22440
	1.71	5.60	25240
	1.88	7.49	27750
P8	1.2	0.04	1590
	3.0	0.18	3980
	5.0	0.39	6630
	7.5	1.05	9940
P10	1.2	0.09	1930
	3.0	0.29	4820
	5.0	0.64	8050
	7.5	1.51	12080

Rearrangement of terms and the substitution of L by the ratio n/q, where q is the number of discs per unit length, results in Equation 15:

$$15) \quad K_L a_{20} = \frac{K_1 K_2}{mC_s} \mu(n\pi(r^2 - r_w^2))^{0.815} \left(\frac{\text{rpm } f}{\upsilon p q}\right)^{1.815}$$

where

$$16) \quad K_1 = 5.7 \times 10^{-8} \text{ (dimensionless)}$$

$$17) \quad K_2 = 35.32 \text{ (mg ft}^3/\text{g}\ell)$$

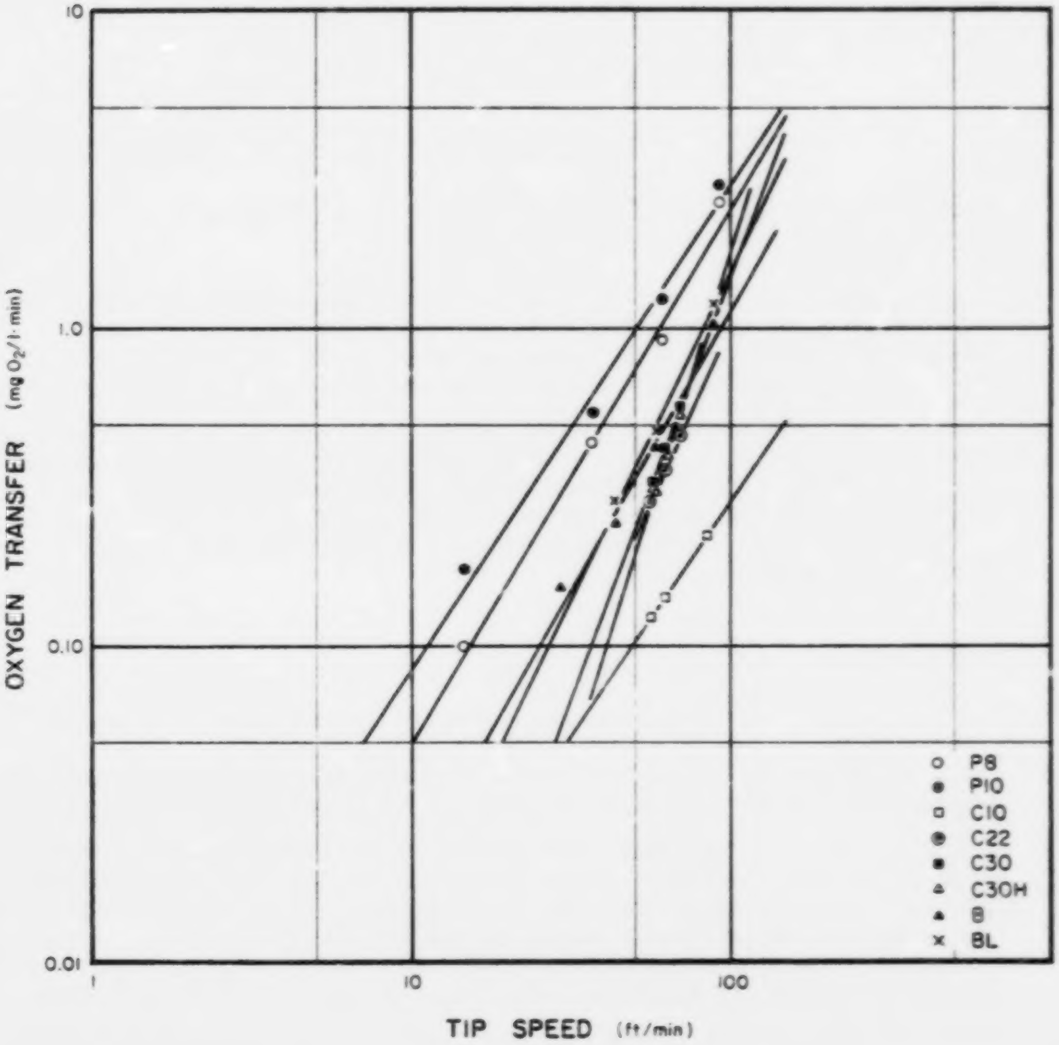
The dimensions of all the parameters in Equation 15 are those given in the nomenclature section.

Equations 13 and 15 can be expected to be valid only for the media design and tank geometry used in these tests. It is speculated that the same general form of the correlation between N_{20} and Re will describe results for other forms of media, but different coefficients will be found. A media design closer to a flat plate should give oxygen transfer results with a lesser dependence upon tank mixing, as described by the Reynolds number. This is presumably due to differences in the water pumping capabilities of media with different designs.

DISCUSSION

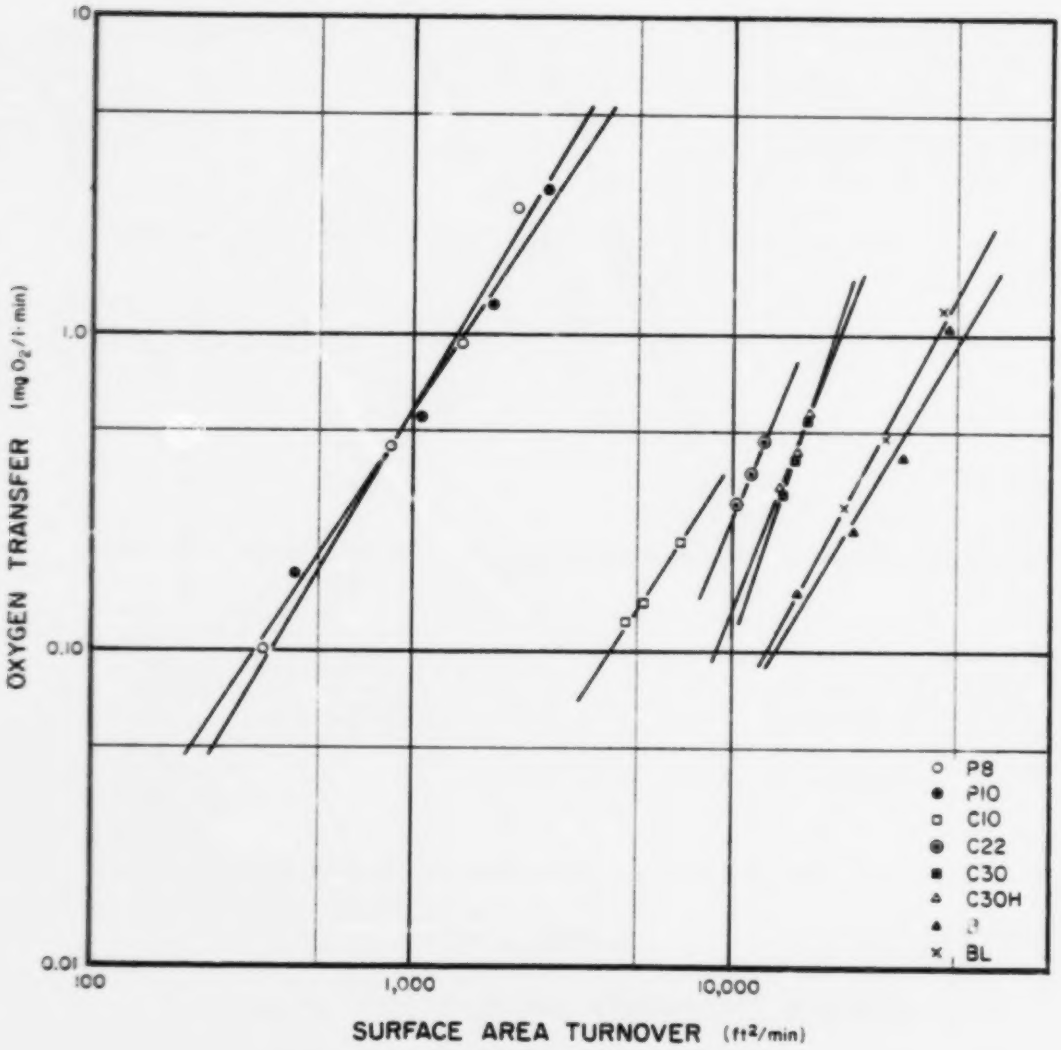
Conventional Scale-up Methods

Figures 4, 5, and 6 show the oxygen transfer results expressed as milligrams oxygen per liter per minute (mg O_2/ℓ min) plotted against the conventional scale-up parameters of tip speed, surface area turnover rate, and rpm, respectively. It is instructive to compare pilot configurations with approximately the same ratio of surface area to tank volume. The pair of pilot configurations which have the closest ratio of surface area to volume are P10 and C30, with ratios of 68.3 and 69.4 ft^2/ft^3, respectively. When comparing the operation of the plants at constant tip speeds, P10 transfers oxygen (mg O_2/ℓ min) at a rate approximately three times that of C30. When operating at constant surface area turnover rates, P10 transfers oxygen at a rate 100 times that of C30. If constant values of rpm are compared,



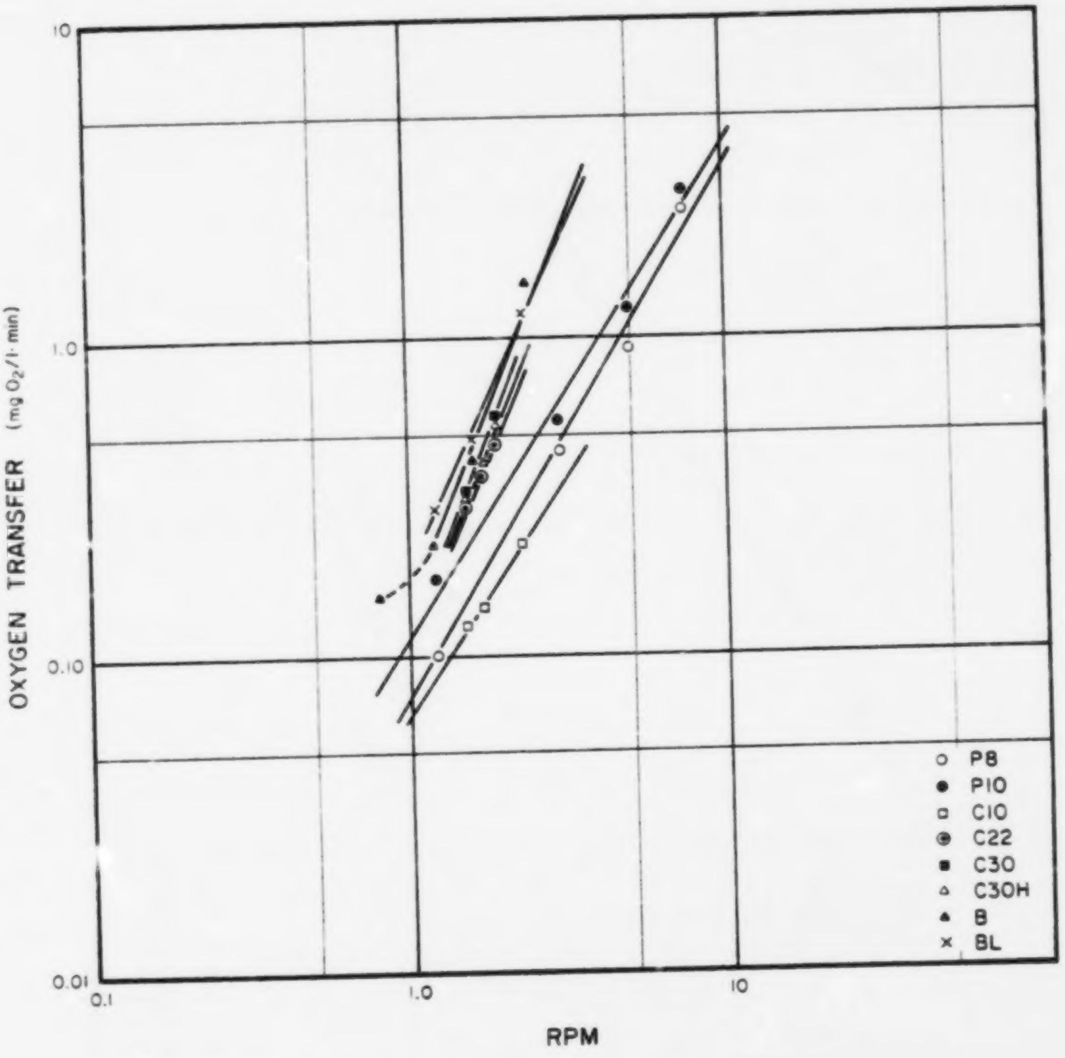
OXYGEN TRANSFER (mg O_2/l· min) versus TIP SPEED (ft/min)

FIGURE 4



OXYGEN TRANSFER (mg O_2/l· min) versus
SURFACE AREA TURNOVER (ft²/min)

FIGURE 5



OXYGEN TRANSFER (mg O_2/l·min) versus ROTATIONAL SPEED (rpm)

FIGURE 6

between the approximate rpm values of 1.5 to 1.9, C30 transfers oxygen at a rate 1.5 to 3 times the rate of P10. From these analyses, it is seen that none of the parameters of tip speed, surface area, or rpm give proper oxygen transfer correlations for scale-up of rotational effects. Scale-up based on tip speed and surface area turnover overestimate the transfer capacity of the large diameter plant. Scale-up by rpm underestimates the transfer capacity of large diameter plants. None of these methods gives as strong a correlation between oxygen transfer and rotational effects as does the relation between N_{20} and Re shown in Figure 3 and Equation 13.

Scale-up By N_{20} and Re

At the present state of knowledge, it is necessary to assume that if two RBC systems transfer oxygen at the same rate (mass of oxygen per unit area per volume) in clean tap water with the media free from biomass, then the oxygen transfer will be the same when the two systems are operated under similar conditions of hydraulic loading, organic loading, and substrate removal. Little data concerning the effects of biomass on the transfer of oxygen to the bulk solution are available to test this assumption.

If the above assumption is made, then Equation 15 may be used to establish the test rpm of a pilot system. If the full scale and pilot systems contain the same ratio of surface area to volume, then it is only necessary to match the oxygen transfer coefficient, K_La_{20}, of the pilot system to that of the full scale RBC. If the surface area to volume ratios in the two systems differ, then the values of $(K_La_{20}V_t/a)$ must be matched. The experimental loading capacity of the pilot RBC is then found. The maximum loading is achieved when the minimum dissolved oxygen limit in the bulk solution can no longer be maintained. Since the first stage of an RBC is subjected to the highest oxygen demand, the critical area for testing the rotational effects on oxygen transfer is in the first stage.

The above statements concerning K_La_{20} and $(K_La_{20}V_t/a)$ were derived by considering mass balances of oxygen and organic substrate in a single stage of an idealized RBC. It is assumed that the single stage is completely mixed. When pilot and full scale RBC systems are compared, it is assumed that constant dissolved oxygen levels must be maintained, and that the rate of oxygen depletion and substrate removal by the biomass is dependent upon the surface area per volume of media available.

The drawback to this scale-up method is that in order to determine the oxygen transfer coefficient of the full scale RBC, it is necessary to assume the tank dimensions and amount of media in one of the stages. This requires that tank design parameters are available from the RBC manufacturer. Some rationale for the design of the full scale plant must be considered, i.e., some estimation of the surface area and staging requirements, prior to the initiation of the pilot testing.

FUTURE RESEARCH NEEDS

In order to use the N_{20} - Re correlation to scale-up oxygen transfer effects, it must be assumed that two RBC systems with equal oxygen transfer capacities in clean water will have the same oxygen transfer capacities when operated under similar conditions of hydraulic and organic loading. This assumption can be tested with side-by-side operation of a pilot and a full scale RBC which have been preset for equal oxygen transfer capacities.

Tests need to be designed to determine where the oxygen demand of the organic loading is exerted, i.e., what fraction of biological oxygen demand is removed at the expense of oxygen available immediately from the atmosphere and what fraction is removed at the expense of oxygen dissolved in the bulk liquid. It may be possible with this information to predetermine the maximum hydraulic and organic loading to an RBC through mass balance considerations.

The N_{20} - Re relation was developed from data taken in tanks which were semicircular in design (Santa Clara and Bensenville) and of angular design (Princeton), such that, the disc edge at all points below the water level was approximately the same distance to the tank wall. The effects on oxygen transfer of square bottom tanks and of multiple shafts in the same tank need to be studied.

The N_{20} - Re relation should be confirmed with 2 foot and 6 foot diameter pilot plants. The ability to rely on 2 foot diameter pilot plants rather than having to use 6 foot systems would increase the available design data by allowing for easier piloting of a waste stream.

The N_{20} - Re relation was not derived from fundamental physical considerations, but was developed empirically. A development of the oxygen transfer capacity of rotating discs from fundamental physical considerations should be undertaken.

The N_{20} - Re relation should be used in conjunction with power tests and organic removal rates to help optimize RBC performance and tank design.

CONCLUSIONS

1) Oxygen transfer data were collected from three RBC pilot plants. Eight test cases using variations in water level, number of discs per tank, and distances between the media and tank bottoms were studied. The results from these tests have been used to develop an empirical correlation between a dimensionless oxygen transfer grouping, N_{20}, and a Reynolds number, Re, incorporating net flow relations in the tanks. The dependence of N_{20} on Re is suspected to be a function of the media design. The closer the design is to a flat plate, the smaller the pumping capacity of the discs and the smaller the dependence of the N_{20} coefficient on the Reynolds term.

2) Conventional methods of analysis of the scale-up of rotational effects between small pilot plants and full scale plants, i.e., tip speed, rpm, and surface area turnover rates, failed to provide a correlation between oxygen transfer data from the eight pilot test cases.

3) A method of scaling the rotational effects between pilot and full-scale RBC systems is suggested where the desired operational rpm of the pilot plant is calculated from tank geometries of the pilot and full scale RBC systems. If the two systems have similar surface area to volume ratios, then the scale-up factor is to equalize the oxygen transfer coefficient, $K_La_{20}(\text{min}^{-1})$. If the surface area to volume ratios are not equal, then the scaling factor is the parameter $(K_La_{20}V_t/a)$, where V_t is the volume of the tank and (a) is the active area of the RBC discs.

NOMENCLATURE

a	active surface area of the RBC (ft^2)
A	area between the RBC media and the tank at the reference cross-section (ft^2)
C	oxygen concentration (mg O_2/l)

C_s	oxygen saturation concentration (mg O_2/l); 9.2 mg/l at 20°C
d	ratio, tank volume/active disc area, V_t/a, (ft)
E	distance from disc edge to tank bottom (ft)
f	void volume; volume of water per volume of disc, 0.935 (dimensionless)
H	water depth of center of tank
K_1	experimental constant, 5.7×10^{-8}, (dimensionless)
K_2	conversion factor, 35.32 (mg x ft^3/ g x l)
K_La	oxygen transfer coefficient (min^{-1})
K_La_{20}	oxygen transfer coefficient corrected to 20°C (min^{-1})
L	length of RBC disc section (ft)
m	area of RBC disc per area of a circle with the same radius, 3.02 ft^2/ft^2, (dimensionless)
n	number of RBC discs (dimensionless)
N_{20}	dimensionless oxygen transfer number
OTR	oxygen transfer rate corrected to 20°C (g O_2/min)
p	wetted perimeter; solid boundaries at reference cross-section (ft)
q	number of discs per length (n/ft) = 12
Q	theoretical pumping rate of the RBC (ft^3/min)
r	RBC disc radius (ft)
r_w	distance from RBC shaft center to waterline (ft)
R	hydraulic radius, A/p, (ft)
Re	Reynolds number (dimensionless)
rpm	rotational speed (min^{-1})
μ(mu)	viscosity (19.19 g/ft min) at 20°C

υ(nu)	kinematic viscosity (65.44 x 10^{-5} ft^2/min) at 20°C
$\bar{v}$	average water velocity (ft/min)
V_d	active volume of RBC discs (ft^3)
V_t	volume of test tank (ft^3)
w	tank width (ft)

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COMPARISON OF FULL SCALE RBC PERFORMANCE WITH DESIGN CRITERIA

by

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I. INTRODUCTION

It is estimated that there are over three hundred wastewater treatment plants using rotating biological contactors since the corrugated surface was first developed and applied in the United States. The majority of the installations in North America have been designed for the removal of BOD and/or ammonia in domestic sewage. In the past decade, design methodology and loading criteria of the RBC process have been defined and consolidated through a great number of pilot studies and field monitoring programs. More recently, Autotrol Corporation has followed the monthly process performance of many RBC treatment facilities in the United States, so that the field operating conditions and treatment efficiency can be updated and analyzed.

By keeping close contact with the operating personnel, not only the appropriate operational measures have been recommended to the plant superintendents, but the process optimization could be realized according to changing environmental conditions. In adition, the field data are regularly examined against design criteria in order to verify the current process design method and to reveal problem areas which affect system performance.

In this report, chronological field data from seven RBC sewage treatment plants will be discussed. These plants have been in operation for one to six years. They are serving the communities of Gladstone, Michigan, Toppenish and Woodland, Washington, Rhinelander, Wisconsin, Georgetown, Kentucky, Caddilac, Michigan, and Fort Knox, Kentucky. The first five were designed for secondary treatment with RBC following either primary clarifiers or fine screens. In the latter two plants, RBC's are used for nitrifying activated sludge effluent (Caddilac), and for removing BOD and ammonia following primary treatment (Fort Knox).

The monthly performance reports from the seven plants for the past one to three years are summarized. The actual performance has been compared with the designed effluent requirements or the NPDES permit conditions.

All the RBC plants being discussed employ mechanical drive systems for rotation and the nominal speed has been 1.6 rpm. The wastewater flows are either normal or parallel to the shafts. Both flat-bottom and contoured tanks are used. Although standard density medium is used in all the plants, latter stages at Rhinelander and Fort Knox are also equipped with High Density media.

II. SECONDARY TREATMENT AFTER GRAVITY PRIMARY CLARIFICATION

The unit operations of the plants at Gladstone, Michigan, Woodland and Toppenish, Washington are shown in Table 1. The original Bio-Surf design flow, surface area, and staging are also tabulated. The wastewater flows in all three plants are parallel to the RBC shafts, and trapezoidal tanks are used. Conventional secondary clarifiers are employed for the removal of RBC effluent suspended solids. Secondary clarifier effluents are chlorinated and discharged. The sludge is anaerobically digested at Gladstone and Toppenish, but aerobically treated at Woodland. The digestor supernatant is recirculated to the head end of the plants. Each shaft is divided by one cross - sectioned baffle, resulting in two stages per shaft.

The monthly average BOD and suspended solids data of the three plants are shown in Table 2, 3 and 4 respectively. The summer and winter performance of each plant is compared with the design requirement.

Gladstone Treatment Plant

This is the first rotating contactor sewage treatment plant built in the U.S. with an EPA construction grant, and has been in operation since March of 1974, for upgrading the original primary treatment. The design and NPDES requirement on BOD and SS is 30 mg/1. Phosphorus removal is achieved by alum and polymer coagulation in two rectangular primary clarifiers.

As shown in Table 2, this plant has not only met the discharge limits on BOD and SS, but has also produced a highly nitrified effluent. The average ammonia nitrogen in the summer months is less than 1.1 mg/1. From April to June 1979, when the average flow was 1.0 MGD, the same as the design flow, the effluent BOD and SS were 15 mg/1 and 17 mg/1 respectively, well below the permit level. Some of the earlier field data and background information of this plant was documented by Antonie (1). In general, the Gladstone plant has been doing very well.

Toppenish Treatment Plant

This 1.3 MGD plant was designed for secondary treatment with discharge requirement of 30 mg/1 and/or eighty-five percent removal of both BOD and suspended solids. The plant was started up in the summer of 1978. In 1979, the hydraulic loading during summer months already surpassed the design level due to infiltration of irrigation water (Table 3). For all the flow and temperature conditions experienced, the process performance has been quite stable. The average effluent BOD and SS are 6 mg/1 and 8-9 mg/1, which account for over ninety-three percent removal. In addition, nitrifiers are believed to dominate on latter stages as indicated by biological slime determinations, although the extent of nitrification awaits to be assessed by ammonia analysis.

Woodland Treatment Plant

This small treatment plant serves a community of twenty-two hundred people. The two-shaft RBC process has been on stream since January of 1975 and requires minimum attention. Each shaft is outfitted with three inter-stage baffles resulting in 4-stage treatment. The monthly average BOD and suspended solids data from August 1978 to July 1979 are summarized in Table 4. The effluent BOD and SS averaged 20 mg/1 and 14 mg/1 for the twelve months. Although the plant is receiving less than the designed hydraulic loading, the waste strength has been substantially higher than domestic waste concentrations and the performance was quite satisfactory despite water temperature declines to 45°F temperature range.

III. SECONDARY TREATMENT AFTER FINE SCREENING

The two plants discussed herein, Rhinelander, Wisconsin and Georgetown, Kentucky, both employ fine screens (Hydrasieves) ahead of RBC process for removing debris from raw wastewater. The flow is perpendicular to the shafts. Raw BOD and suspended solids loads are essentially unchanged by the screening and thus, solids loadings are somewhat higher than at plants having primary clarifiers. There are two RBC tanks in both plants. The Rhinelander plant has flat-bottom tanks with vertical interstage baffles, whereas the Georgetown plant has a contoured tank baffling configuration.

The unit operations and design information of the two installations are shown in Table 5. The lower design hydraulic loading at Rhinelander is due to the more stringent design effluent requirement and lower water temperature in the winter. Process performance at these plants has been very satisfactory and relatively unaffected by the additonal solid loading in the contactors, as the plants continue to produce effluent quality below permit requirements.

Rhinelander Treatment Plant

The average flow from July 1978 to August of 1979 is seven percent higher than the 1.9 MGD design flow as shown in Table 6. The influent and effluent BOD are almost identical with the projected levels. The influent suspended solids is 11 mg/l higher than the designed 160 mg/l. It is reasoned that the nature and concentration of the influent SS and the higher overflow rate in the final clarifiers result in the average 20 mg/l effluent SS, which is 4 mg/l higher than the expected design figure. However, the effluent has been consistently in compliance with the 30 mg/l discharge permit conditions, and a few of the operational problems such as malfunctioning grit chamber, faulty digestor operation, and insufficinet ventilation in the RBC building were corrected in the past year.

Georgetown Treatment Plant

The RBC process was started in June, 1975. The wastewater is of medium strength in BOD and SS. A summary of recent monthly data is shown in Table 7. The performance has been quite stable as indicated by the consistent BOD and SS removal of over 90% for the twelve months surveyed. The average final effluent BOD and SS are both 16 mg/l, well below the 30 mg/l discharge limit. Again, the lack of primary solid removal does not affect the wastewater treatment efficiency of the rotating contactors, although no consideration was taken for the extra solids load in the original design.

IV NITRIFICATION OF ACTIVATED SLUDGE, EFFLUENT

The advanced treatment plant at Cadillac, Michigan, employs an activated sludge system with $FeCl_3$ addition for BOD and phosphorus removal. The Bio-Surf process is used for nitrifying secondary clarifier effluent with final polishing provided by sand filters. (A flow diagram of this installation is shown in Figure 1). The plant was designed for 2 MGD and the RBC section was started up in March of 1976. Eight shafts with a total surface area of 7.6 x 10^5 sqaure feet are installed in two parallel flat-bottom tanks. The flow is normal to the shafts through slots at the bottom of interstage baffles. The nitrification process has been operated in a four-stage mode.

The NPDES requirement on effluent ammonia nitrogen is 1.5 mg/l from June to October. During the summer of 1976, the effluent ammonia nitrogen slightly exceeded the limit. It was later revealed by a thorough tracer study funded by the Autotrol Corporation, that the RBC nitrifying capacity was penalized by short circuiting along the bottom of the tank. Subsequently, a simple adjustment was made to the baffle dimensions to improve residence time distribution in June of 1977. The effluent ammonia nitrogen since that time has been below the NPDES permit shown in Table 8.

Since nitrification is only required during the summer, the biological contactors are stopped during the winter. By doing so, savings in energy and O & M labor can be achieved without affecting other sections of the treatment plant. Usually, the rotating contactors at Cadillac are restarted in April so that active nitrifying biomass can be developed to meet stringent effluent ammonia requirements from June 1st. Dissolved oxygen in RBC tanks varies from 3.5 to 4.0 mg/l, and alkalinity declines 8.0 mg/l per part per million of ammonia nitrogen oxidized. The pH in RBC effluent is about 7.6 during the summer months, when the influent pH is 7.8. Thus, the nitrification proceeds at a rather favorable pH range.

V BOD AND AMMONIA REMOVAL FOLLOWING PRIMARY SETTLING

The 6 MGD sewage treatment plant at Fort Knox, Kentucky was designed for combined removal of BOD and ammonia. After the conventional primary treatment, wastewater is distributed to six flat-bottom RBC tanks. There are three standard media and three High Density media shafts in each bay, with verticle interstage baffles between adjacent shafts. Thus, the RBC process is operated in six-stage mode. The High Density media on stages 4 through 6 is supposed to support the relatively thin nitrifying biomass.

The thirty-six Bio-Surf shafts were started in November of 1977. The performance data from November of 1978 to October of 1979 are shown with the effluent requirements in Table 9. Both the effluent BOD and the suspended solids have met the design quality. The winter ammonia nitrogen also met the discharge limit of 5 mg/l, but the average summer effluent ammonia is 2.7 mg/l, exceeding the required 2.0 mg/l standard.

A field examination of the biomass on the last three stages reveals that the apparent autotrophic biofilm is extremely thin and the color is darker than the average nitrifiers on the contactor surface. It is suspected that the unique biofilm appearance and the inadequate nitrifying activity might have been caused by the sub-neutral pH in RBC tanks, especially toward the latter stages where pH as low as 6.5 has been detected. The range of the pH of RBC effluent is also shown in Table 9.

Earlier studies have shown that the nitrifiers either suspended growth or attached thin film would be severely inhibited when the pH drops below

6.8 (2) (3). A subsequent full scale kinetic study conducted at the Autotrol Corporation also demonstrates that the nitrification rate declines from 0.31 lb. of ammonia nitrogen removed per thousand square feet per day at pH 7.0 to 0.17 lb/1000^{2}ft. day at pH 6.5. At Fort Knox, the critical pH in the nitrification stages is the result of insufficient buffer capacity of the wastewater. Although the average influent total alkalinity is about 160 mg/1, the pH tends to decrease below 7.0 when the first forty percent of the alkalinity is consumed by the nitric acid produced. The amount of alkalinity consumed is equivalent to the removal of about 9 mg/1 ammonia nitrogen. At this point, the nitrification rate would slow down significantly at ammonia concentration less than 6 mg/1 due to both the unfavorable pH and the decreasing ammonia concentration. As one would expect, nitrification kinetics is first order with respect to ammonia when its level falls below 5 mg/1 and removal proceeds at slower rates.

VI COMPARISON OF BIO-SURF PERFORMANCE WITH DESIGN CURVES

The process design of the RBC treatment plant discussed above were based upon Autotrol's criteria, which have resulted from a great deal of effort in piloting, monitoring full scale operation and kinetic modeling (4) (5). The major design curves for both BOD and ammonia removal are attached as Appendix I and II. Since the fixed-film biological contactors absorb and oxidize the impurities in the soluble forms, the general design practice has been the derivation of the required surface area from the influent soluble BOD (and/or ammonia), and the desired effluent soluble BOD ammonia oxidation (Appendix III and IV). If soluble BOD data is not available, it is estimated by subtracting the suspended portion according to the empirical formula shown in Appendix V.

For ease of comparison, the hydraulic loading, the calculated influent soluble BOD, the field effluent BOD and ammonia are shown in contrast with the predicted effluent quality in Table 10. Among the seven plants, Gladstone, Toppenish, Rhinelander, Georgetown and Cadillac are producing better effluent than predicted. Woodland's effluent BOD has also been meeting the NPDES permit requirement, however, it shows some deviation from the predicted level which might be due to unusual sludge handling practices. This plant awaits to be further assessed when the hydraulic loading approaches 3 to 4 gpd/ft^2 in the future.

At Fort Knox, the effluent BOD is essentially the same as predicted. The actual ammonia nitrogen, however, is higher than the calculated figure. As discussed earlier, this discrepency is reasoned to be caused by the unfavorable pH in the nitrification section. It is concluded that the nitrifying capacity of the equipment is penalized by the acidic environment.

VII CONCLUSIONS

1. All the seven plants surveyed have met the NPDES permit requirements on effluent BOD and suspended solids.

2. The RBC plant treating activated sludge effluent has met both the NPDES limit and design predictions on effluent ammonia

nitrogen.

3. Five out of the six plants designed for BOD removal have performed up to Autotrol's process design criteria. The only exception is the underloaded Woodland treatment plant, which awaits further assessment when hydraulic loading approaches design level.

4. The data suggests that the elimination of primary settling of suspended solids does not affect the BOD removal capacity of the RBC process.

5. The Fort Knox treatment plant, designed for combined BOD and ammonia removal, is meeting the effluent BOD, SS, and the winter ammonia requirement. The average summer time effluent ammonia is 0.7 mg/1 higher than the required 2.0 mg/1. Investigations point to retardation effects on nitrification at the sub-neutral pH values observed at this installation.

VIII REFERENCES

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2. Process Design Manual for Nitrogen Control

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3. Huang, C. S., and N. E. Hopson, "Temperature and pH Effect on the Biological Nitrification Process". Presented at the Annual Winter Meeting, New York Water Pollution Control Assoc., N.Y.C., January, 1974.

4. Famularo, J., J. A. Mueller and T. Mulligan, "Application of Mass Transfer to Rotating Biological Contactors" JWPCF 50: 653-671, 1978.

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FIGURE 1

PROCESS FLOW DIAGRAM OF CADILLAC WASTEWATER TREATMENT PLANT

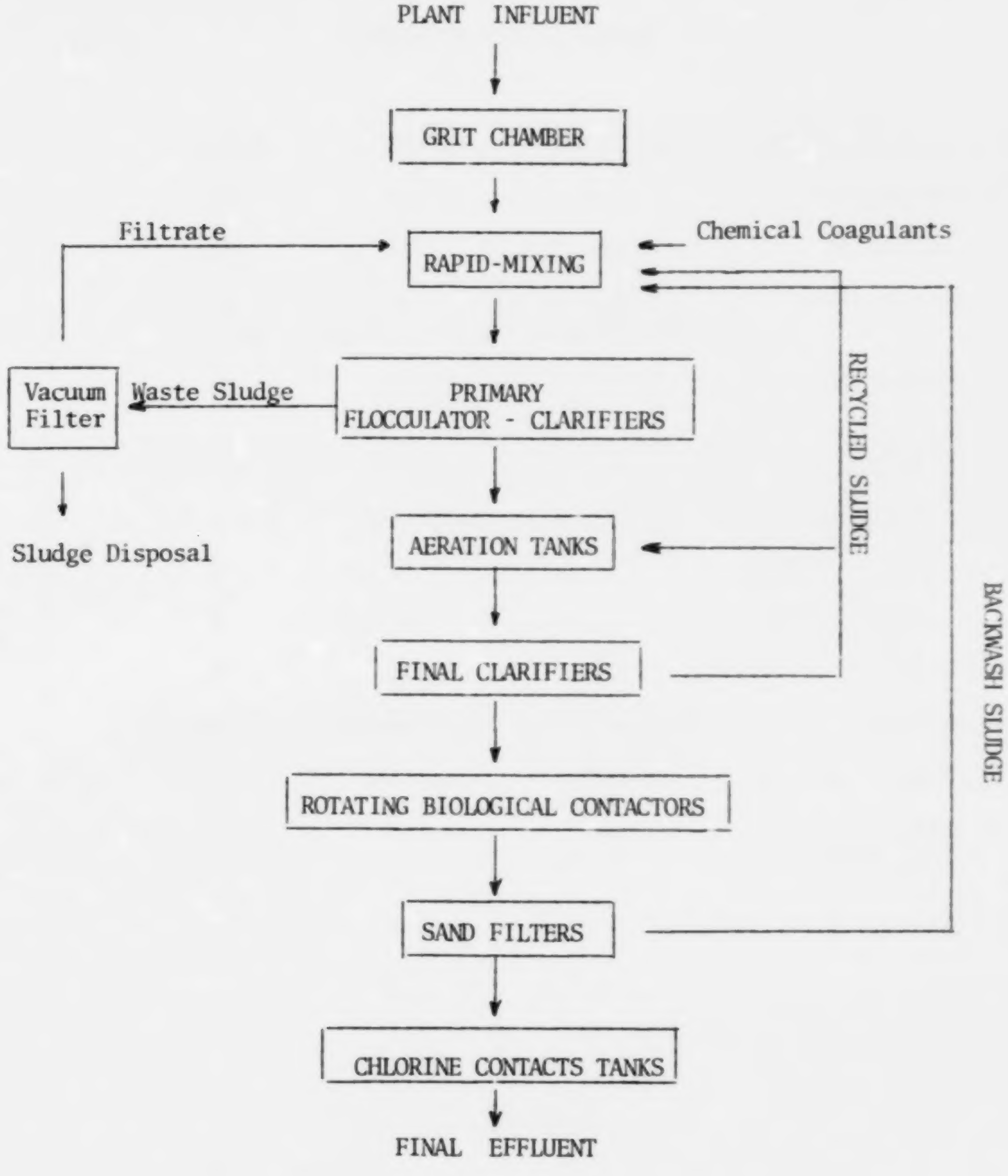

PLANT INFLUENT
GRIT CHAMBER
Filtrate
RAPID-MIXING
Chemical Coagulants
Vacuum Filter
Waste Sludge
PRIMARY FLOCCULATOR - CLARIFIERS
RECYCLED SLUDGE
Sludge Disposal
AERATION TANKS
BACKWASH SLUDGE
FINAL CLARIFIERS
ROTATING BIOLOGICAL CONTACTORS
SAND FILTERS
CHLORINE CONTACTS TANKS
FINAL EFFLUENT

TABLE 1

UNIT OPERATIONS AND DESIGN DATA OF GLADSTONE, TOPPENISH AND WOODLAND SEWAGE TREATMENT PLANT

UNIT OPERATION	PLANT GLADSTONE	 TOPPENISH	 WOODLAND
1. Bar Screen	Yes	Yes	Yes
2. Comminutor	Yes	Yes	Yes
3. Grit Chamber	Yes (non-aerated)	Yes (non-aerated)	Yes (aerated)
4. Primary Clarifier	Yes	Yes	Yes
5. Bio-Surf			
Design Flow, MGD	1.0	1.30	0.56
Design SA ft^2	5.2×10^5	5.9×10^5	1.4×10^5
No. of Stages	6	4	4
No. of Shafts	6	6	2
6. Secondary Clarifier	Yes	Yes	Yes
7. Disinfection	Chlorination	Chlorination	Chlorination
8. Sludge Digestion	Anaerobic	Anaerobic	Aerobic
9. Sludge Dewatering vs Disposal	Land	Land	Land

TABLE 2

SUMMARY OF MONTHLY DATA AT

GLADSTONE, MICHIGAN

PERIOD	FLOW		BOD, mg/1		SS (mg/1)		NH_3 (mg/1)	
	MGD	°F	RBC INFL.	FINAL EFFL.	RBC INFL.	FINAL EFFL.	RBC INFL.	FINAL EFFL.
January '78	0.59	49	112	5	70	17	15.8	4.0
February '78	0.48	47	112	6	76	20	-	3.7
March '78	0.54	46	92	12	63	16	16.4	3.3
April '78	0.80	47	70	13	49	18	10.1	2.6
May '78	0.75	53	78	8	52	17	13.0	1.6
June '78	0.59	60	79	6	60	14	15.8	0.7
July '78	0.55	64	73	7	59	14	20.1	1.3
August '78	0.57	67	107	7	70	16	-	0.6
September '78	0.77	65	80	6	79	16	12.6	1.0
October '78	0.74	61	91	5	77	18	-	<1.1
November '78	0.52	57	103	4	71	17	17.5	1.2
December '78	0.44	51	110	7	67	18	-	4.2
January '79	0.41	47	86	13	66	15	16.8	3.0
February '79	0.42	46	80	13	55	15	18.6	3.7
March '79	0.65	46	60	12	48	16	15.9	4.5
April '79	1.04	46	49	16	51	18	7.8	2.7
May '79	1.00	50	41	14	33	18	7.6	<2.0
June '79	0.95	55	46	14	42	16	8.9	<1.0
July '79	0.77	61	48	10	48	18	10.9	<1.0
August '79	0.86	63	44	10	43	18	7.4	<1.0
Summer Average May - October	0.76	60	69	9	56	16	12.0	<1.1
Winter Average November - April	0.59	48	87	10	62	17	14.9	3.3

TABLE 2

SUMMARY OF MONTHLY DATA AT

GLADSTONE, MICHIGAN
(Continued)

	FLOW MGD	°F	BOD, mg/1 FINAL EFFL.	SS (mg/1) FINAL EFFL.
Design Condition	1.0		30	30
NPDES Permit			30	30

Estimated RBC Influent Soluble BOD = RBC Influent Total BOD - (0.6) (RBC Influent SS)

TABLE 3

SUMMARY OF MONTHLY DATA AT

TOPPENISH, WASHINGTON

PERIOD	FLOW		BOD_T mg/1		S.S. mg/1	
1979	MGD	°F	RAW	FINAL EFFL.	RAW	FINAL EFFL.
January	0.73	54	177	7	104	9
February	0.73	55	153	6	159	6
March	0.72	57	146	5	178	16
April	0.90	57	118	*	136	*
May	1.33	62	183	8	183	6
June	1.50	64	126	4	189	5
July	1.57	64	86	8	162	11
August	1.61	65	109	6	125	7
September	1.34	66	100	4	124	8
October	1.05	64	114	4	169	13
November	0.89	60	175	5	149	5
Summer Average: May to October	1.40	64	120	6	159	8
Winter Average: November to April	0.79	57	154	6	145	9
Deisgn ** Condition	1.3			30		30
NPDES Permit**	1.3			30		30

Estimated RBC influent soluble BOD = (o.75) (Raw BOD) -(0.6) (0.4) (Raw SS)

* Sampler malfunctioned, available data not included in average
** And/or 85% reduction of raw BOD and SS

TABLE 4

SUMMARY OF MONTHLY DATA

WOODLAND, WASHINGTON

	FLOW		BOD_T, mg/1		SS, mg/1	
PERIOD	MGD	°F	RAW	EFFL.	RAW	EFFL.
August '78	0.207	66	243	20	296	16
September '78	0.225	64	247	19	287	10
October '78	0.223	61	249	23	430	15
November '78	0.214	56	205	18	358	19
December '78	0.249	50	200	22	184	15
January '79	0.247	48	271	20	235	11
February '79	0.310	45	213	20	211	18
March '79	0.302	47	176	22	225	18
April '79	0.307	48	178	18	174	8
May '79	0.289	50	162	19	170	11
June '79	0.282	52	160	20	177	14
July '79	0.272	56	103	22	203	14
Summer Average	0.250	58	194	20	260	13
Winter Average	0.272	49	207	20	231	15
Design Condition	0.56			30		30
NPDES Permit				30		30

TABLE 5

UNIT OPERATIONS AND DESIGN DATA OF PLANTS USING SCREENS FOR INITIAL TREATMENT

UNIT OPERATION	PLANT RHINELANDER, WISCONSIN	GEORGETOWN, KENTUCKY
1. Bar Screen	Yes	Yes
2. Grit Chamber	Yes	No
3. Fine Screen	Yes	Yes
4. Bio-Surf (RBC)		
Design flow, MGD	1.9	3.0
Design Surface, ft.^{2}	11×10^5	8×10^5
Number of Stages	5	4
Number of Shafts	10	8
5. Secondary Clarifier	Yes	Yes
6. Disinfection	Chlorination	Chlorination
7. Sludge Digestion	Anaerobic	Anaerobic
8. Sludge Dewatering	Belt Press	Sludge Bed

TABLE 6

SUMMARY OF MONTHLY DATA AT

RHINELANDER, WISCONSIN

	FLOW		BOD, mg/l		SS, mg/l	
PERIOD	MGD	°F	RAW	EFFL.	RAW	EFFL.
July '78	2.08		122	13	163	23
August '78	2.27		152	14	191	25
September '78	2.98		187	16	171	23
October '78	3.07		154	18	131	21
November '78	1.85		144	18	166	22
December '78	1.45		119	14	167	19
January '79	1.34		153	18	207	24
February '79	1.44		144	15	164	22
March '79	1.75		130	15	154	17
April '79	2.12	46	116	12	168	23
May '79	2.09		120	10	132	15
June '79	2.36		191	16	243	22
July '79	1.87	59	155	12	181	11
August '79	1.75		121	11	162	14
Average:	2.03		143	14	171	20
Design Condition	1.9		145	15	160	16
NPDES Permit				30		30

Estimated RBC influent soluble BOD = Raw BOD - (0.6) (Raw SS)

TABLE 7

SUMMARY OF MONTHLY DATA AT

GEORGETOWN, KENTUCKY

PERIOD	FLOW, MGD	BOD_T (mg/1)		SS (mg/1)	
		RAW	FINAL EFFL.	RAW	FINAL EFFL.
July '78	1.08	202	15	202	17
August '78	1.24	212	14	207	14
September '78	1.11	206	15	216	16
October '78	1.19	201	14	202	12
November '78	1.36	204	14	205	15
December '78	1.87	225	16	195	14
January '78	1.69	214	14	204	12
February '79	1.62	224	17	204	16
March '79	1.31	206	16	205	17
April '79	1.65	213	18	207	18
May '79	1.06	239	20	216	18
June '79	1.00	205	16	201	17
Annual Average	1.35	213	16	205	16
December to April '79	1.63	216	16	203	15
Design Condition	3.0		30		30
NPDES Permit	3.0		30		30

Estimated RBC Influent Soluble BOD = Raw BOD - (0.6) (Raw SS)

TABLE 8

SUMMARY OF MONTHLY DATA AT

CADILLAC, MICHIGAN

PERIOD	FLOW, MGD.	WASTEWATER TEMPERATURE °F	NH_3-N, mg/1		
			RBC INFL.	RBC EFFL.	FINAL EFFL.
November '77	1.36	60	10.2	1.2	
December '77	1.31	55	16.7	2.6	
January '78	1.42	53	16.3	2.2	2.0
February '78	1.47	52	15.5	2.4	2.3
March '78	1.46	53	15.0	2.7	2.5
May, '78	1.36	58	4.3	1.0	0.7
June '78	1.44	65	9.6	2.0	1.8
July '78	1.36	68	6.6	0.6	0.6
August '78	1.41	69	8.6	1.5	1.5
September '78	1.42	67	8.8	0.9	0.8
October '78	1.52	63	6.8	0.4	0.2
June '79	1.62	63	13.1	1.4	1.0
July '79	1.61	66	8.9	0.7	0.4
August '79	1.61	67	8.6	0.8	0.6
September '79	1.50	66	5.9	0.6	
October '79	1.51	63	11.0	1.5	1.2
Summer Average June to October	1.50	66	8.5	1.0	0.86
Summer NPDES Requirement					1.5

TABLE 9

SUMMARY OF MONTHLY DATA AT

FORT KNOX, KENTUCKY

PERIOD	FLOW MGD	°F	RBC EFFL. pH	BOD mg/1 RBC INFL.	BOD mg/1 FINAL EFFL.	SS mg/1 RBC INFL.	SS mg/1 FINAL EFFL.	NH_3-N, mg/1 FINAL EFFL.
November '78	3.59	70		87	10	78	20	4.5
December '78	4.32	68		77	13	77	13	1.7
January '79	4.16	68		112	16	150	13	2.0
February '79	4.19	68		134	22	117	17	2.6
March '79	3.75	69		108	3	130	24	3.9
April '79	4.48	70		85	10	97	23	4.2
May '79	3.54	70		99	9	107	19	2.9
June '79	4.00	71		112	9	190	25	2.2
July '79	4.19	71		112	8	167	23	1.9
August '79	3.90	71	6.7-7.0	60	12	175	12	2.7
September '79	3.94	70	6.5-6.8	71	9	85	9	4.5
October '79	3.51	70	6.8-7.0	140	7	100	28	2.3
Average Summer	3.85	70		99	9	137	19	2.7
Average Winter	4.08	69		100	14	108	18	3.2
Design and Permit)	Summer				10		30	2
Condition)	Winter				20		30	5

Estimated RBC Influent Soluble BOD = RBC Influent Total BOD - (0.5) (RBC Influent SS)

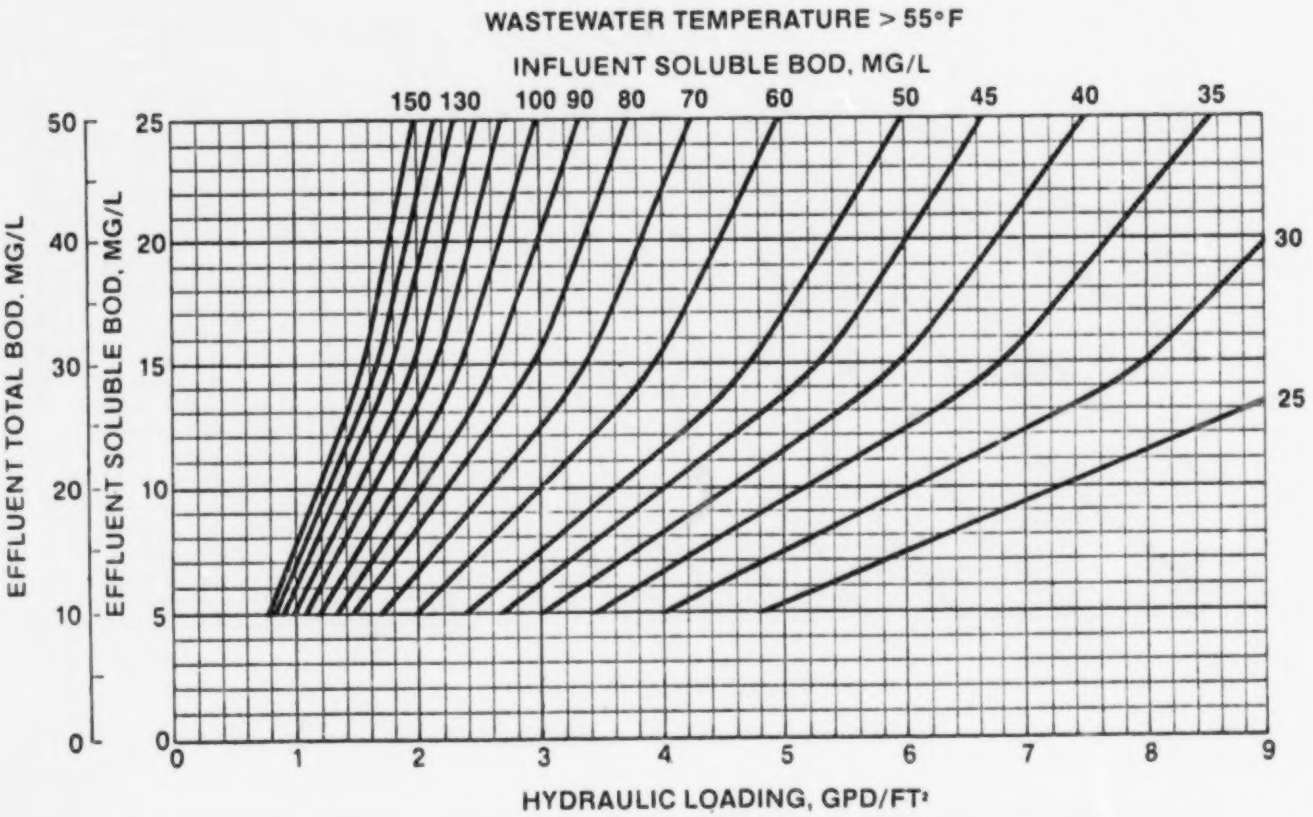
TABLE 10

COMPARISON OF FIELD PERFORMANCE WITH AUTOTROL'S PREDICTION

INSTALLATION	HYDRAULIC LOADING gpd/ft^2	SOLUBLE BOD, mg/1			NH3-N, mg/1	
		INFLUENT CALCULATED	EFFLUENT PREDICTED	EFFLUENT ACTUAL	EFFLUENT PREDICTED	EFFLUENT ACTUAL
Gladstone						
Summer	1.4	35	5	4	1.0	<1.1
Winter	1.1	50	5	5		
Toppenish						
Summer	2.4	52	5	3		
Winter	1.3	80	5	3		
Woodland						
Summer	0.45	84	5	10		
Winter	0.49	100	5	10		
Rhinelander						
Annual	1.8	40	5	7		
Sept. Oct. '78	2.7	80	12	8		
Georgetown						
Annual	1.7	90	7	8		
Dec. 78 - Apr. '79	2.0	94	10	8		
Cadillac						
Summer	2.0				1.0	1.0
Fort Knox						
Summer	0.86	31	5	4	1.0	2.7
Winter	0.91	46	5	7	1.0	3.2

APPENDIX 1

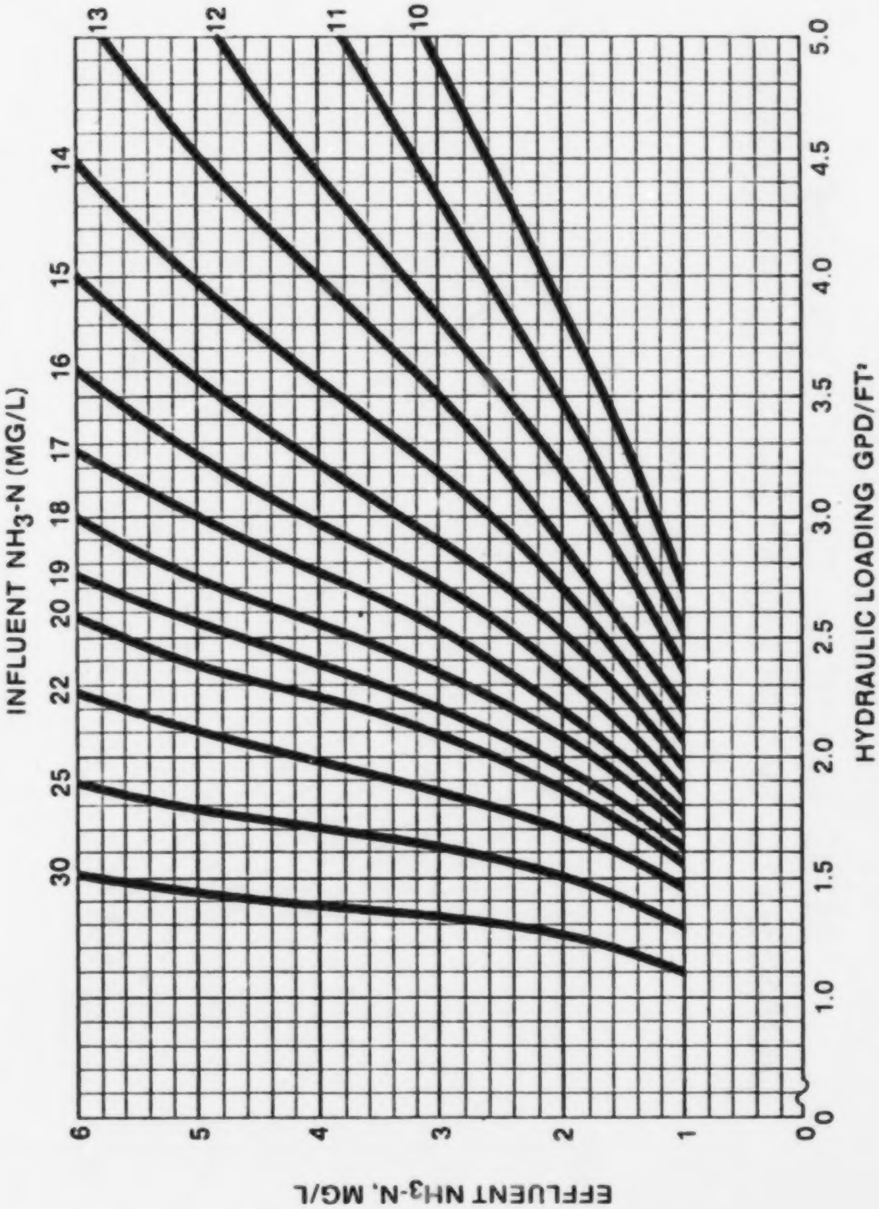
BIO-SURF PROCESS
MECHANICAL DRIVE SYSTEM
DOMESTIC WASTEWATER BOD REMOVAL



APPENDIX 2

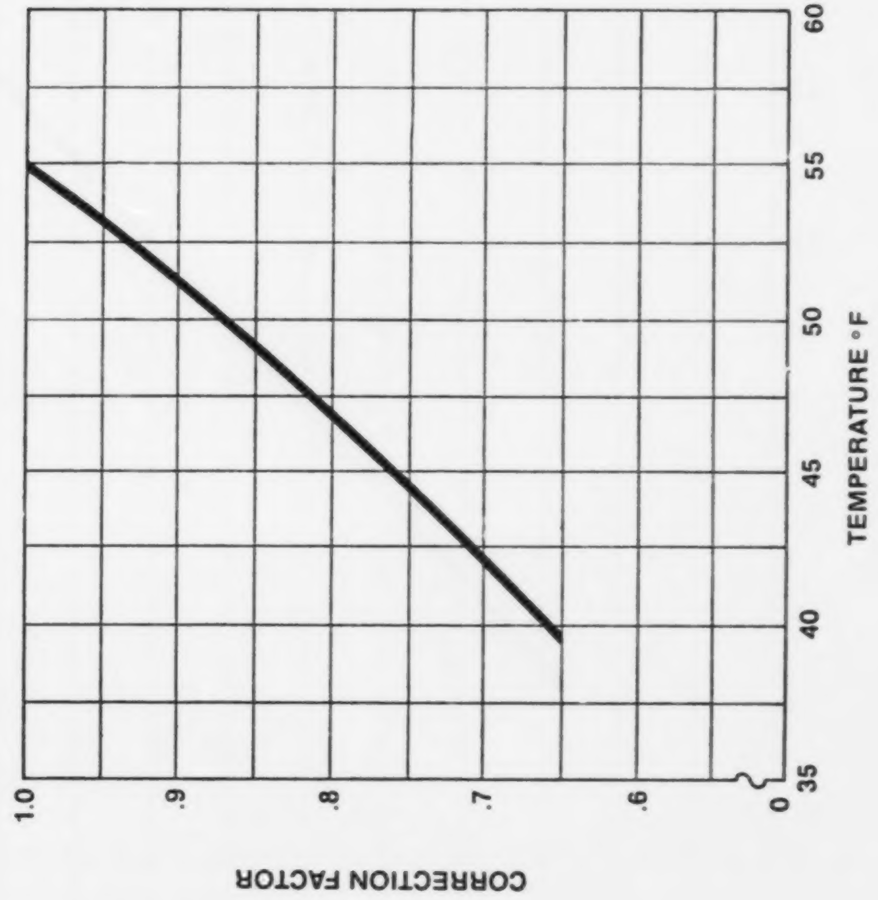
NITRIFICATION OF DOMESTIC WASTEWATER

WASTEWATER TEMPERATURE > 55°F



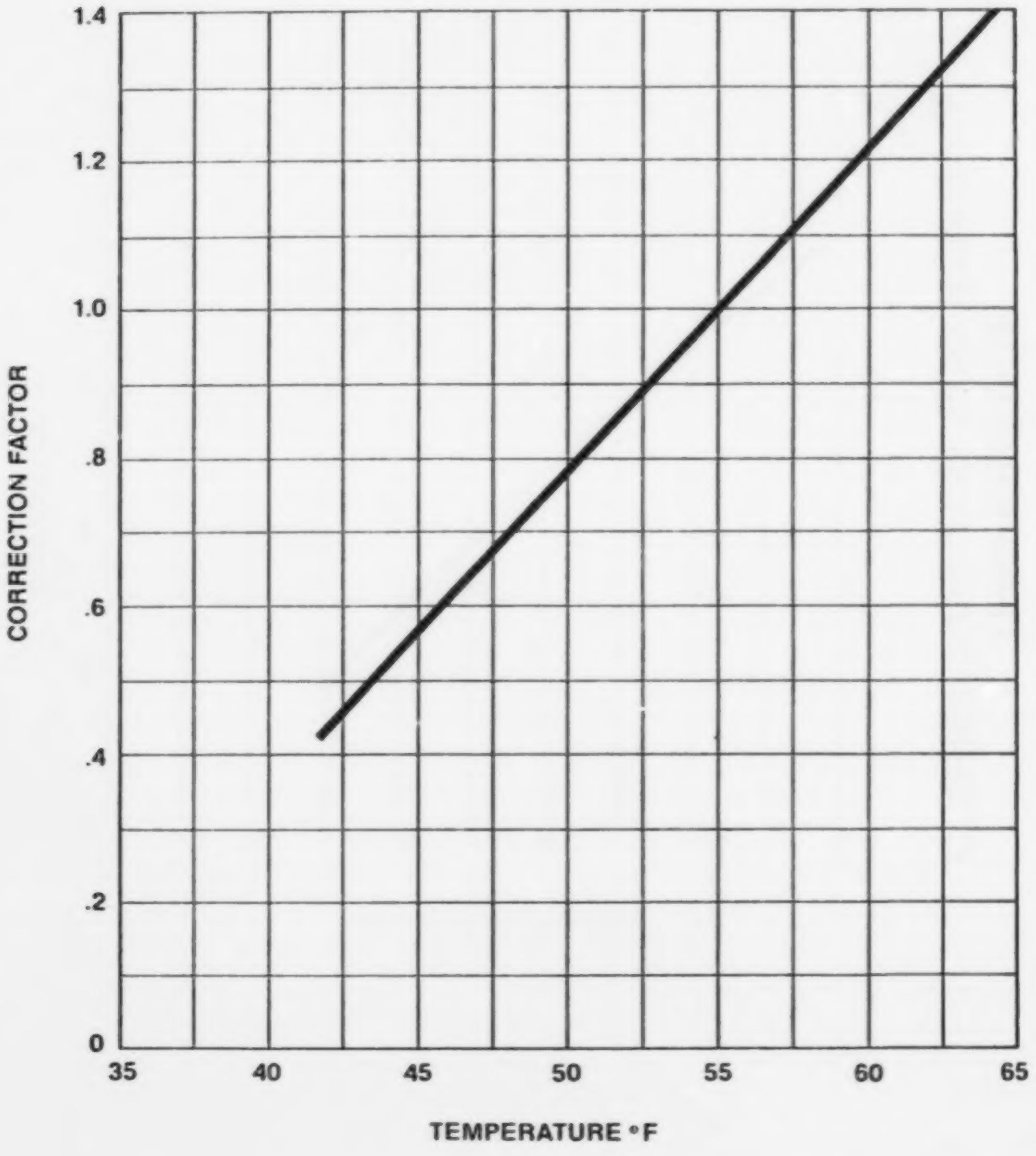
APPENDIX 3

TEMPERATURE CORRECTION FOR BOD REMOVAL



APPENDIX 4

TEMPERATURE CORRECTION FOR NITRIFICATION



APPENDIX 5

SOLUBLE BOD

Soluble BOD is determined by conducting a standard BOD test on wastewater that has passed through the fiberglass mat filter used for a standard suspended solids test. This gives results comparable to a millipore filter but the test is faster and easier to perform. Although a workable test with consistent results, this procedure is not a measure of materials in "true solution" because some colloidal materials will pass through the fiberglass filter. For conducting soluble BOD tests refer to the procedure in the Appendix of this manual.

For design purposes, soluble BOD test data on the wastewater to be treated is, by far, the most useful information. If the wastewater flow exists, whether it is a raw waste or primary effluent, it is a simple matter to run BOD tests on the filtrate from the standard suspended solids test. For purposes of preliminary studies, initial cost estimates, or if the wastewater does not exist, it is possible to estimate the soluble BOD concentration from historical data on total BOD and suspended solids, or by assuming appropriate values for these parameters. Soluble BOD can be estimated as follows:

Soluble BOD = Total BOD — Suspended BOD

Suspended BOD = K x Suspended Solids

Soluble BOD = Total BOD — K x Suspended Solids

where

K = 0.5 to 0.7 for most domestic wastes

= 0.5 on raw sewage where suspended solids values are significantly higher than the total BOD values e.g., BOD = 200 mg/l and SS = 250 mg/l (typical of strictly residential area)

= 0.6 on raw sewage where suspended solids values are about the same as the total BOD values (typical of municipalities with commercial and industrial discharges)

= 0.6 on primary treatment effluents

= 0.5 on secondary treatment effluents

FIRST USA AIR DRIVE RBC UNITS
OPERATIONAL EXPERIENCE AND PERFORMANCE
INDIAN CREEK WASTEWATER TREATMENT PLANT
CINCINNATI (CLEVES), OHIO

By

M. M. Schirtzinger, President
M. M. Schirtzinger & Associates, Limited
Consulting Environmental Engineers
Chillicothe, Ohio

PLANT DESCRIPTION

The Indian Creek Wastewater Treatment Plant provides secondary treatment for an average daily flow of 0.5 million gallons per day (mgd) and receives peak flows up to 1.7 mgd of domestic sewage. Six air-driven rotary biological contactors (RBCs) (Aero-Surf units supplied by the Autotrol Corporation) are preceeded by grit removal and two 40-foot diameter primary clarifiers and are followed by two 40-foot diameter secondary clarifiers. Secondary sludge is returned to the primary clarifiers and combined with primary sludge. Sludge is removed from the primary clarifier to a sludge holding tank and is trucked away to another plant location for ultimate disposal.

The Indian Creek Wastewater Treatment Plant (WWTP) is located on the banks of the Ohio River at the foot of a monument to William Henry Harrison and is bordered by park land on two sides. The plant was designed by M. M. Schirtzinger & Associates, Limited (MMSAL) and is operated by the Cincinnati Metropolitan Sewer District (MSD).

The plant receives domestic sewage from the communities of Cleves and North Bend and the southwestern part of Hamilton County. Flow through the plant is by gravity. Raw waste strength is low in biological oxygen demand (BOD) averaging from 100 to 130 milligrams per liter (mg/l), and no industrial waste is received for treatment.

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FIRST USA AIR DRIVE RBC

The plant was originally designed using mechanically driven 20-foot shafts in three stages. Just prior to the beginning of construction, the plant was modified to use the air drive system because of the additional reliability which the air drive units would provide, the power savings which could be expected, and because the flood elevation made protection of the mechanical drive units difficult.

Two sets of three single-stage 20-foot shaft length RBC units are arranged for parallel flow. Each RBC has 80,000 square feet of surface area (a total of 480,000 square feet). Each set of three RBCs is in a common tank with dividing baffles to provide three separate stages. Ammonia reduction is not required at this plant. Air is supplied by a single Lamson blower with a 40 horsepower (H.P.) driver.

The original mechanically driven RBC units would have been equipped with 7.5 H.P. drive units requiring a total connected horsepower of 45. The RBC air drive blower is provided with a 40 H.P. motor; however, it is throttled to use 25 H.P. or 4.16 H.P. per drive. This results in a savings in the daily power use. However, the precise savings in power is not known since the power draw for the mechanical drives is unknown and because a portion of the air supply from this blower is used to aerate the feed well of the primary clarifiers.

ADVANTAGES OF THE AIR DRIVE UNITS

The advantages of the air drive units are not limited just to power savings as other positive benefits are available which include: (1) individual speed regulation of each stage, (2) fewer moving parts and electrical equipment which are subject to corrosion, (3) augmentation of the bioreaction due to the air supply, (4) the use of air for other unit processes, (5) the elimination of effluent reaeration, (6) no decrease in efficiency during wintertime operating temperatures, and (7) safety. These advantages are more fully discussed as follows:

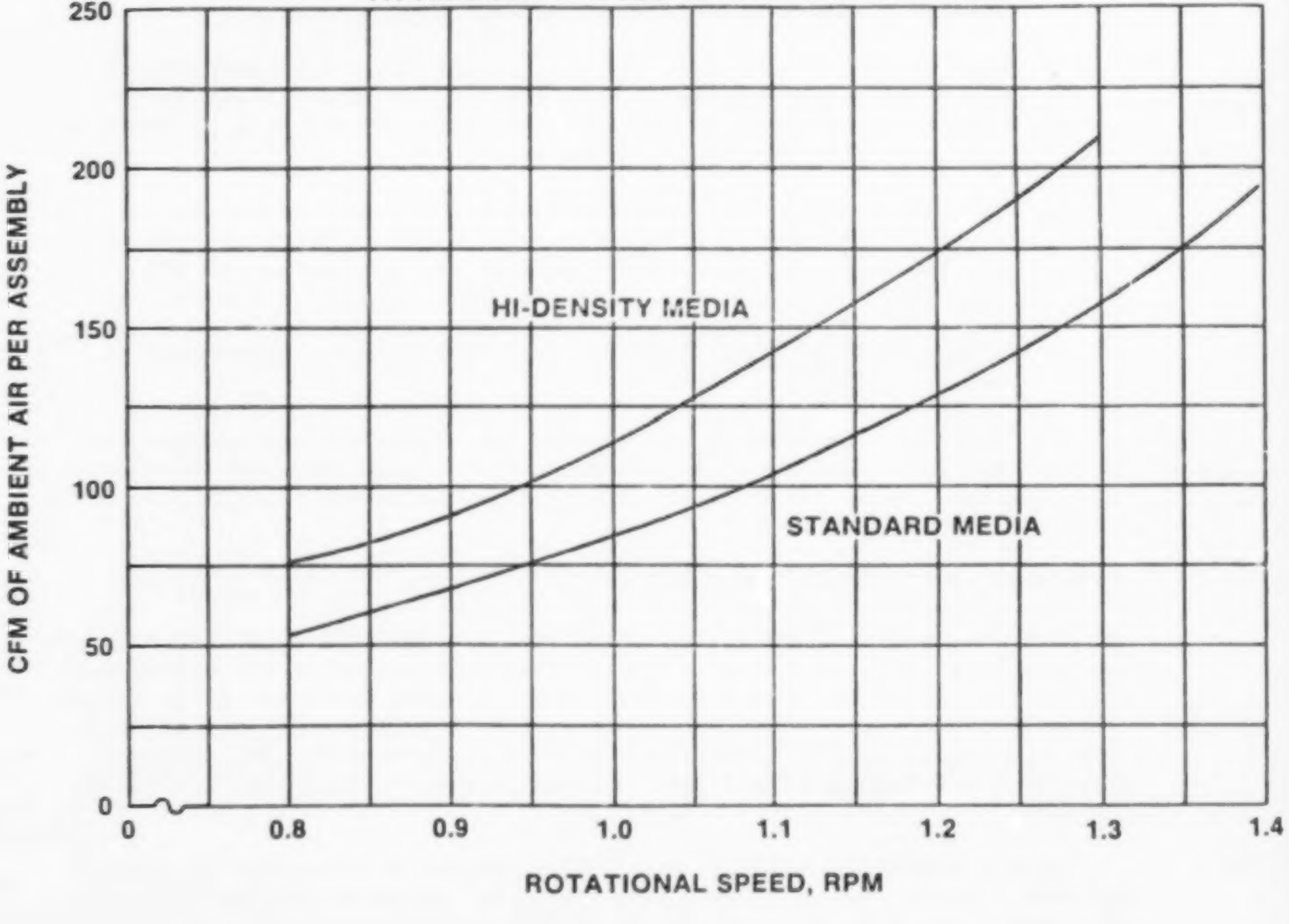
1. Because the air flow to each stage can be regulated by a throttling butterfly valve, the revolutions per minute (rpm) of each stage can be individually adjusted. Thus, each stage rpm can be proportional to the BOD_5 applied. The first stage would have the highest rpm. The optimum speed of the units ideally would be 1.3 rpm, 1.1 rpm, and 1.0 rpm. For units with low BOD_5 load, air flow and power can be reduced and speed can be as low as 0.8 rpm. The relationship between speed of rotation and quantity of air is shown on Figure 1.

2. Since there are no mechanical or electrical parts inside the RBC enclosure, corrosion and oxidation of chain drives, gear reducers,

FIGURE I

AIR REQUIREMENTS

BASED ON 25-FOOT MEDIA ASSEMBLIES
AT AMBIENT PRESSURE AND TEMPERATURE



motors, and electrical connections is eliminated. The air header is corrosion resistant PVC pipe. Valves are located outside the enclosure above ground and are easy to adjust and maintain.

3. The use of air augments the unit processes by providing the attached growth system with a surplus of oxygen. For high strength wastes and heavy biomass, increased air is required as shown on Figure 2. At the Indian Creek plant, dissolved oxygen (DO) increases through each stage for a total DO increase of 1 to 3 mg/l.

4. Because air is available, it can be used for other unit processes in the plant such as to aerate grit chambers and feed wells of the primary clarifier; and all the air can come from a single air blower unit. This results in capital cost savings because the number of motors, starters, wire, etc. can be reduced.

5. Because of the air added for driving the RBC units, effluent reaeration is not required. Sufficient DO is added to the waste flow to maintain DO in the final effluent above 5 mg/l as required in the Indian Creek plant NPDES permit.

6. In cold climates, the covered RBC units are not subjected to freezing or a decrease in biological activity as a result of cold ambient air temperatures. Tests with ambient air temperature of 0°C show that there is an increase of 1°C in the wastewater from the primary clarifier effluent to the discharge of the final RBC stage. This added heat comes from the heat of compression generated in the air blower.

7. Since all lubrication and adjustments can take place outside the RBC enclosure, personnel do not have to enter into the RBC enclosures and thus safety hazards are eliminated.

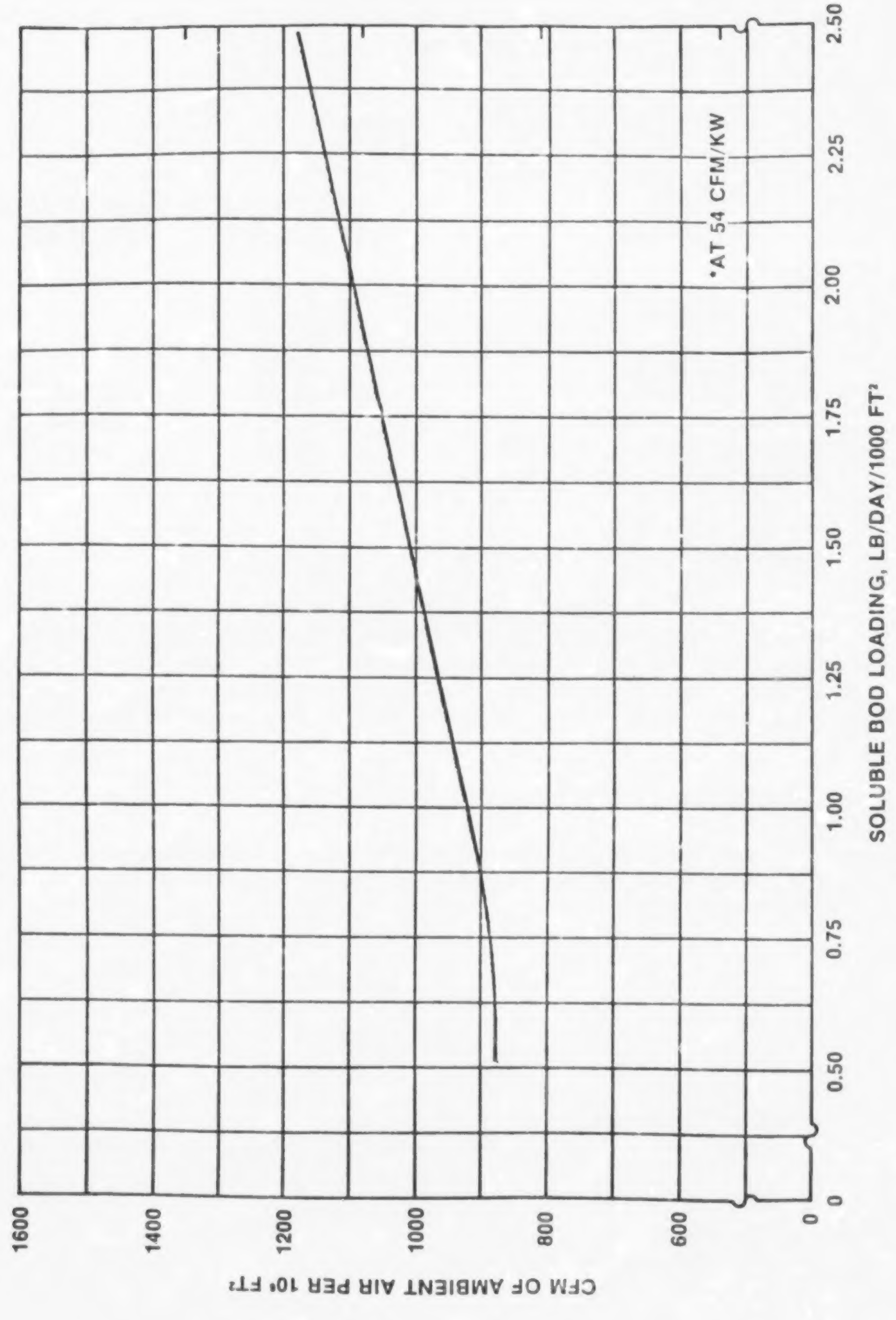
DISADVANTAGES OF THE AIR DRIVE UNITS

The only disadvantage to air drives would occur on a long term shutdown of the RBC units whereby an unbalanced biological growth on the wetted portion of the RBC units becomes so heavy that the unit would not again rotate with only the buoyancy provided by the air drive system. If such a situation occurred, it would be necessary to totally dewater the RBC units so that the biomass sloughed off and the shaft was again in balance to resume rotation.

This disadvantage would be avoided by the use of a backup air blower and with standby power. These backup units are normally provided for in most plants and would be required for continuity of operation on Environmental Protection Agency financed projects.

FIGURE 2

RBC AIR REQUIREMENTS



PLANT PERFORMANCE

Average performance of the Indian Creek plant for the reduction of BOD_5, suspended solids, ammonia nitrogen, and nitrate nitrogen removal as well as other parameters is shown on Table I. Data listed is averaged for the various loading rates shown.

TABLE I

Parameter	Percent Reduced at Loading Rates		
	0.65 gpd/sf	1.04 gpd/sf	2.08 gpd/sf
BOD_5	97.0%	89.5%	83.3%
BOD_5 (Soluble)	95.4%	-	-
COD_5	82.0%	-	-
N_k (Soluble)	91.2%	-	-
NH_3H	100.0%	91.6%	90.2%
Organic N	42.9%	54.0%	50.2%
Suspended Solids (Total)	94.8%	90.0%	85.5%
Suspended Solids (Volatile)	97.1%	-	-

COMPARISON WITH DESIGN DATA

Actual performance of the plant compared to data used for design purposes shows that the Autotrol design data is reliable for air drive units. Comparing the BOD_5 loading and effluent obtained from the operation at the Indian Creek plant shows excellent correlation with the design data presented in Figure 3 for soluble BOD_5 loading in the range of 40 to 70 mg/1.

CORRELATION OF OTHER DATA

Correlation with other operational parameters at a loading rate of 0.65 gpd/sf is shown in Table II for the shafts designated as North (N-1) and South (S-1), etc.

TABLE II

	North Shafts			South Shafts		
	N-1	N-2	N-3	S-1	S-2	S-3
Biofilm Thickness, inches	0.067	0.013	0.011	0.040*	0.011*	0.005*
Effluent Soluble BOD_5, mg/1 (Influent: 33 mg/1)	12	4	2	6	3	1

FIGURE 3

DOMESTIC WASTEWATER BOD REMOVAL

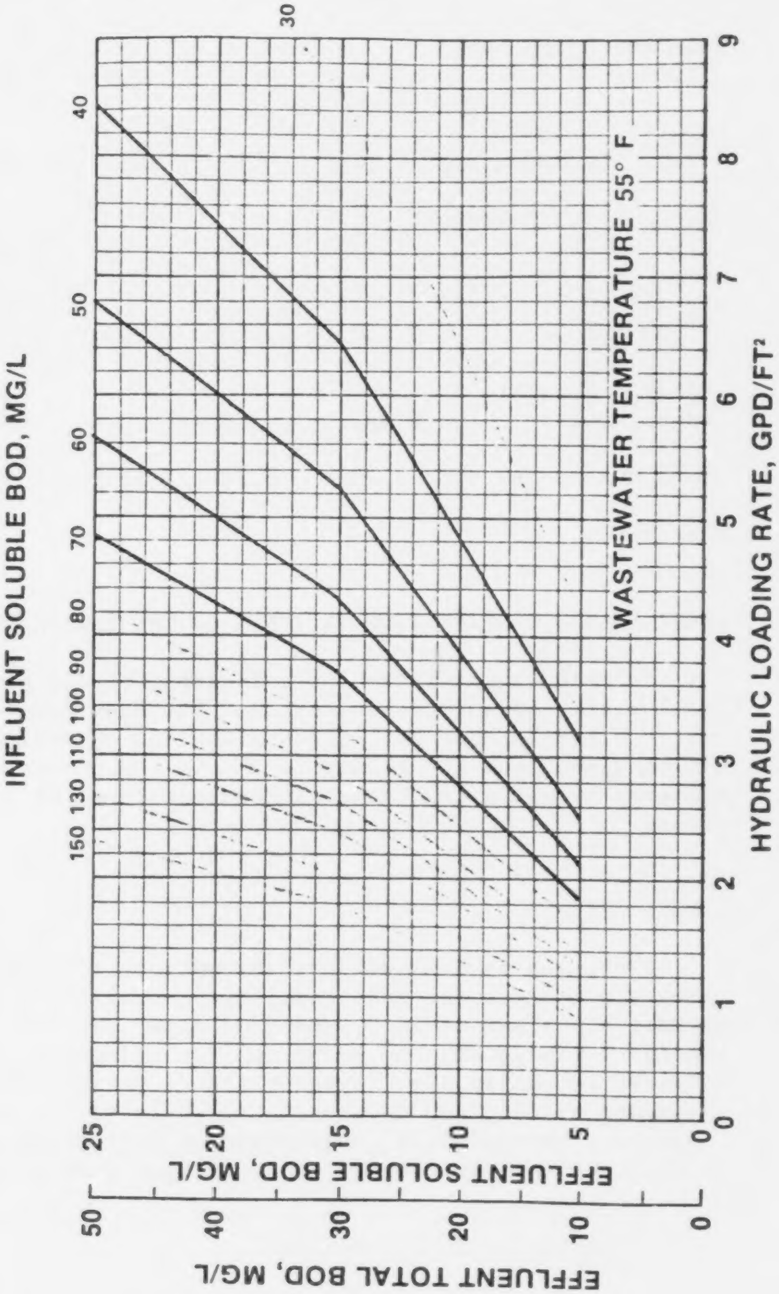


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TABLE II
(Continued)

	North Shafts			South Shafts		
	N-1	N-2	N-3	S-1	S-2	S-3
Applied Carbon, lbs/ 1000 ft^2	0.54	0.20	0.07	0.54	0.10	0.05
Specific Removal Rate, Carbon	0.34	0.13	0.04	0.44	0.05	0.03
NH_3-N, mg/l (Influent: 15.4 mg/l)	11.1	1.6	0.0	2.8	0.0	0.0
Applied HN_3-N, lbs/ 1000 ft^2	0.250	0.180	0.026	0.250	0.045	0.000
Specific Removal Rate, NH_3-N	0.070	0.154	0.026	0.205	0.045	0.000
Shaft RPM	1.18	1.20	1.36	1.54	1.28	1.36

*Estimated

OTHER ENERGY CONSERVATION FEATURES OF THE INDIAN CREEK PLANT

In order to maximize the efficiency of all power use at the Indian Creek plant, advantage was taken of the heat generated by the RBC blower units. This is used for building heating by having the furnace intake air obtained from the blower room. In the summer, room ventilation is achieved by using the blower intake air from the laboratory and pump room areas.

REMOTE OPERATION

The plant is designed for automatic unattended operation. No adjustments are made to the rotational speed of the RBC units, and all sludge withdrawal is performed by pumps on timer controls. Intermittent recycle of secondary sludge and the withdrawal of primary sludge to holding tanks is achieved automatically. Plant effluent parameters such as DO, temperature, suspended solids, and pH are continuously monitored and transmitted to the Cincinnati Metropolitan Sewer District Mill Creek Plant. Chlorine residual is monitored and controlled with a closed loop chlorine residual analyzer.

RELIABILITY

Since its initial operation in January 1978, the plant has operated without mechanical failures nor interruption of flow for any reason. Blocking of air lines or difficulties with air-driven units has not been experienced nor has the sludge handling equipment malfunctioned during that period of time. The only difficulties experienced to date have been with remote monitoring sensors for DO, pH, and suspended solids.

SUMMARY AND CONCLUSIONS

Based upon the experienced to date at the Indian Creek plant, air driven RBC units are highly reliable providing the designed level of treatment efficiency at low power input rates. Air drives have no disadvantages and have numerous advantages over mechanically driven units. These advantages include (1) individual speed regulation of each stage, (2) fewer moving parts and electrical equipment which are subject to corrosion, (3) augmentation of the bioreaction due to the air supply, (4) the use of air for other unit processes, (5) the elimination of effluent reaeration, (6) no decrease in efficiency during wintertime operating temperatures, and (7) safety.

Acknowledgements

The author hereby acknowledges the assistance and cooperation of the Cincinnati Metropolitan Sewer District, Richard A. Vanderhoof, Director, in providing data and cooperating in making plant changes.

The graphs and other data presented in this paper were by courtesy of the Autotrol Corporation.

DESIGN AND OPERATION OF TWO ROTATING BIOLOGICAL CONTACTOR PLANTS AT FUNDY NATIONAL PARK, NEW BRUNSWICK, CANADA

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INTRODUCTION

Rotating biological contactor (RBC) systems were selected to upgrade existing wastewater treatment plants (septic tanks) treating wastewater from two campgrounds at Fundy National Park, New Brunswick, Canada, during the summer months. As part of the design and selection process, an investigation was carried out on the performance of RBC equipment in general and RBC equipment manufactured by a Canadian manufacturer in particular. Results of this investigation are presented in the paper. The plants

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became operational in the summer of 1979. The design features, the operational controls and the performance of these plants are described in the paper.

TREATMENT ALTERNATIVES CONSIDERED AND THE CHOICE

The two treatment plants, one at Headquarters campground and the other at Point Wolfe Campground, were to be designed to treat flows of 18 600 gpd and 26 000 gpd respectively and analysis results indicated that the septic tank effluent was relatively weak in strength (BOD less than 100 mg/1) on occasions sampled.

Package plants incorporating suspended growth biological systems like extended aeration, contact stabilization were not considered suitable to treat a weak wastewater with fluctuating flows as it would be difficult to develop optimum MLSS for operation and these plants would require more skilled operation. Though a well-designed physical-chemical treatment system would be able to accomplish the desired results, such was not considered because it would require considerable supervision as well as skilled operation. It was felt that fixed growth systems would normally be acceptable treatment systems in this case. However, trickling filter systems, because of possible odour and fly problems and relatively lower BOD removal efficiencies, were not recommended.

Rotating biological contactor systems were aesthetically more acceptable; these were found to be more efficient, comparatively, in respect of BOD removal and were found to be ideal for summer camp applications treating weak wastewaters and fluctuating flows. RBC systems were found to require minimum start-up time, viz., only one week, and the operation of these plants is quite simple. Results of a study done by Sack and Phillips (1) on the evaluation of a RBC system for summer camp application indicated that these systems are especially suitable in such applications. Based on the above analysis, it was proposed to use RBC systems to upgrade the septic tank effluent in the two locations.

SUMMARY RESULTS OF GENERAL INVESTIGATION OF RBC EQUIPMENT

A general investigation of RBC equipment, and their performance was conducted. Results of the investigation are outlined below.

Autotrol Corporation, Milwaukee, Wisconsin pioneered the manufacture of RBC equipment in North America. RBC equipment are presently manufactured by others listed below, in addition to Autotrol.

1. Environmental Systems Division,
   Geo. A. Hormel and Company
   Coon Rapids, Minnesota

2. Tait/Bio-Shafts Inc.,
   Oconomowoc, Wisconsin

3. Envirodise Corporation (Clow)
   Beacon, New York

4. CMS Equipment Limited
   Mississauga, Ontario

There are a few more manufacturers of RBC units but these were not evaluated.

It was learned that RBC units installed as of August, 1978 in the states of Wisconsin and Maine and in the Province of Alberta were manufactured by Autotrol. There were six RBC plants in the Province of Alberta, all reported to be performing satisfactorily with only a few minor problems of adjustment and with no problem involving either shaft malfunction or failure. All four RBC plants either built or under construction in Maine were Autrotrol units and one was operating for ayear with no significant mechanical problems (subsequently, in the fall of 1979 2 out of 5 shafts failed at the Fort Fairfield plant). All the RBC units in operation in Wisconsin were Autotrol units. Other RBC units are presently included in the specifications as these are reported to be comparable in design. It was understood that the Wisconsin Environmental Department was reasonably happy with Autotrol plants (mechanical aspects and secondary treatment requirements). A perusal of design bulletins and drawings of the RBC manufacturers showed that these were very similar with only minor variations.

This investigation revealed that generally some problems with the shaft surfaced in some locations and necessary rectifications were carried out by the manufacturer as part of development work. It was considered prudent to check the design of these shafts in detail during shop drawing stage and to take sufficient care during installation to minimize any problems.

SUITABILITY OF ROTORDISK UNITS

The suitability of Rotordisk units and related equipment manufactured by CMS Equipment for use at Fundy National Park was specifically assessed as this equipment was found to be very competitive. Information on the performance of seven installations obtained through telephone enquiry, as part of the investigation, revealed that three units exhibited mechanical faults -- two related to the shaft and one related to the gear reduction unit. Two of these units were the first and the third installations of CMS Equipment. Enquiries revealed that the fabrication of the Rotordisk units had improved with time and experience gained in the early installations. The process performance was generally satisfactory in all cases. It was recommended, based on this enquiry as well as review of the general specifications and design, the CMS RotorDisk be considered as an approved equal. Based on this recommendation, and the competitive nature of their quotes (approximately 30% economical) Parks Canada selected RotorDisk units for use at Fundy National Park. During a review of shop drawings, the drive shaft sizes were increased to the next higher sizes to be safe against fatigue failure.

DESCRIPTION OF PLANTS, FLOW PATTERN AND DESIGN CRITERIA

Each treatment system consists of an existing septic tank, a rotating biological contactor unit with clarifier and a chlorination system. An arrangement is provided in both the systems to recycle clarifier effluent through the system during periods of low flow. Sludge collected from the clarifier and scum removed from either the recycle tank or the chlorine contact tank are returned to the septic tank. Flow diagrams for the plants are presented in Figures 1 and 2. Design criteria for the two plants are shown in Table 1.

TABLE 1 - DESIGN CRITERIA

Particulars	Headquarters Campground Treatment Plant	Point Wolfe Campground Treatment Plant
Design population	532	740
Total BOD assuming BOD contribution of 0.05 lb/cap/day, lb/day	27	37
Maximum daily flow, gpd	18 600	26 000
Peak flow, gpd	45 600	74 400
BOD of raw wastewater, mg/l	172	172
SS of raw wastewater, mg/l	165	173
Expected BOD removal efficiency of septic tank %	30	30
BOD of septic tank effluent, mg/l	120	120
Expected BOD removal efficiency in RBC (at 12-8^{0}C) %	84	87
Expected BOD of plant effluent, mg/l	19	16
Expected SS of plant effluent, mg/l	10-20	10-20
Septic Tank		
Capacity of existing septic tank, gallons	14 100	30 000
Septic zone capacity, gallons	9 300	13 000
Equalization zone capacity, gallons	4 800	17 000
Expected sludge accumulation in an operating season, gallons	5 200	5 600
Surface area of tank, ft^2	270	450

TABLE 1 - DEISGN CRITERIA (CONT'D)

Particulars	Headquarters Campground Treatment Plant	Point Wolfe Campground Treatment Plant
RBC-RotorDisk Unit		
Capacity of feed mechanism, gpm	20	27
RBC media diameter, ft	7	10
RBC media area, ft^2	15 000	24 000
Design hydraulic loading (average flow) gpd/ft^2	1.2	1.1
Design BOD loading (average) $lb/d/1000\ ft^2$	1.2	1.1
Secondary Clarifier		
Surface area, ft^2	75	110
Overflow rate (average flow) gpd/ft^2	250	240
Overflow rate (peak flow) gpd/ft^2	610	680
Chlorine Contact Tank		
Capacity, gallons	1 300	800
Detention time, (average flow) minutes	100	45
(peak flow) minutes	41	15
Chlorine dosage, mg/l (maximum)	10	10

OPERATION AND CONTROL OF TREATMENT PLANTS

Sack and Phillips (1) recommended that recycle of sewage from the final clarifier during periods of low flow should be made an automatic function in order to keep the biomass in an optimum condition. The two RBC plants at Fundy National Park incorporate this special design feature of effluent recycle to septic tanks at times of low flow as well as sludge and scum return to the septic tanks. These aspects are elaborated in this section.

HEADQUARTERS PLANT OPERATION AND CONTROL

The pumps deliver the septic tank effluent to the RBC module. The average pumping rate the RBC module is controlled in the range of 12 to 20 gpm, preferably closer to 12 gpm, through (1) throttling the valve on the by-pass line which relieves back to the sump or (2) through throttling both the by-pass valve and the valve on the pump discharge line. These have to be made by manual adjustments, and once adjusted, they are marked/recorded and left undisturbed generally. The recycle pump is controlled by the flow rate over the V-notch weir within a specific pre-set range.

The pumping rate for the recycle pump is controlled such that a maximum flow of 10 gpm is obtained; this is achieved by throttling the two valves (one by-pass and the other on discharge) similar to the case of primary pumps. The recycle pump will operate when the flow rate at the v-notch weir is less than 5 gpm as sensed by the flow recorder. The pump will cease operation when the flow increases to 8 gpm. Within the flow rate of 0 to 8 gpm leaving the treatment plant, the recycle pump will be controlled by float switches.

Sludge from the clarifier is automatically transferred to the sludge and scum tank through a pipe from the revolving clarifier mechanism. Another pipe is used to transfer the scum from the recycle tank to the sludge and scum tank, when the scum collector is operated manually. The sludge and scum pumps deliver the sludge and scum to the inlet end of the septic tank.

Measuring, recording and totalizing of the effluent flow rate is provided utilizing a bubbler tube immersed in a $22\frac{1}{2}^{0}$ V-notch weir located in the recycle tank. The back pressure produced by the bubbler system (directly proportional to the liquid level flowing over the weir) is sensed by the recorder (housed in a separate enclosure) which extracts the weir formula for recording and totalizing. An electronic signal, proportional to the flow, is transmitted to pace the chlorinator. Also located within the recorder instrument case are signal contacts, set to open and close at adjustable flow rates for operation of the recycle pump.

POINT WOLFE PLANT OPERATION AND CONTROL

The septic tank effluent flows by gravity to the RBC module in this case. The recycle pump operation and control is similar to the one described for Headquarters Plant. The v-notch for flow measurement is placed in the chlorine contact tank in the present case. The scum is also collected from the chlorine contact tank. Sludge and scum pumping system and the control system are similar to the ones at Headquarters Plant.

BUILDING FOR RBC PLANTS

The rotating biological contactor should be covered or enclosed. The enclosure serves two main functions. Operation in low temperature climates requires a housing to limit wastewater temperature reduction or heat loss. Also, exposure of the rotating surfaces to direct sunlight and to rainfall should be prevented to stop the growth of algae on the outer surfaces, and to prevent rain wash-off, respectively. The enclosure also helps to prevent deterioration of the plastic medium by ultraviolet rays from sunlight.

Wood frame buildings are provided to house the RBC units, chlorination system, recycle system, flow measurement units and electrical controls. The structures not only provide weather protection for the units and their associated controls, but also help to maintain aesthetic appearance of the Park, with these buildings conforming to others in the area. No heating is provided for the buildings. Air within the enclosure is at a temperature approximately equal to that of the wastewater. At lower ambient temperatures, the high humidity within the buildings will usually result in condensation on the walls and ceiling. Condensation can be minimized by the provision of forced ventilation of the enclosure, drawing fresh air through louvres and removing the moisture-laden air using an exhaust fan. Corrosion-resistant materials are specified for use within the buildings. Interior wood surfaces have been Pentox-treated to resist rot. Reapplication of Pentox will be required over the years.

The building floors are sloped toward peripheral channels at the exterior walls for drainage during washing down. Hose bibs connected to the Park potable water system, are provided for this purpose. The floors of the buildings are insulated from frost eliminating the need to disconnect piping during the winter shut down of the plants.

PERFORMANCE OF THE RBC PLANTS

Headquarters Plant

The plant became operational in the summer of 1979. The plant was inspected in late August to observe its performance. The biomass growth on the RBC was peculiar in that it was quite excessive on the first module, with very much less on the next module and almost nothing on the last module. In the first module, about 25% of the gaps between the woven media sheets were filled with biofilm (sludge). Although there were malodours in the room, foul odours were evident when the sludge deposits on the RBC were disturbed. The biomass was grey-brown and had a furry or hair-like appearance. It was believed that a leak in the septic tank to the wet well contributing septic liquor and sludge was causing the peculiar development of biomass on the RBC unit. The RBC influent has a strong unpleasant odour suggesting a strong wastewater and this was borne out by the results of analytical tests given below.

Sample	BOD mg/l	SS mg/l
RBC influent	165	47
RBC effluent	48	35
Final clarifier effluent	30	15
Efficiency of removal	82%	68%

Point Wolfe Plant

This plant also became operational in the summer of 1979. The performance was observed in late August. The biological growth on the RBC units appeared healthy and normal, with biomass decreasing progressively from the first module to the fourth module. The thickness of the biomass on the face plates of the first module was approximately 0.12 in., decreasing to 0.04 in on the fourth unit. The interstices of the woven mesh media were 70-80% occupied in the first module, with a light film on the solid portion of the media in the last module. The results of analytical tests performed at this plant are shown below.

Sample	BOD mg/l	SS mg/l
RBC influent	50	22
RBC effluent	32	33
Final clarifier effluent	10	9
Efficiency of removal	80%	59%

The percentage removals in respect of BOD and SS are in the same range for both the RBC plants. Preliminary results thus indicate that the performance of these plants is generally good. However the final effluent quality in the case of Headquarters Plant is not as good due to the much stronger wastewater applied. It is believed that the final effluent BOD will fall below 20 mg/l once the leak in the septic tank and recycle problems at Headquarters Plant are set right.

A comprehensive study of the performance of both these plants would be desirable.

REFERENCE

1. Sack, W.A., and Phillips, S.A., "Evaluation of the Bio-Disc Treatment Process for Summer Camp Application" EPA-670/2-73-022, U S Environmental Protection Agency, Washington, D.C., August 1973.

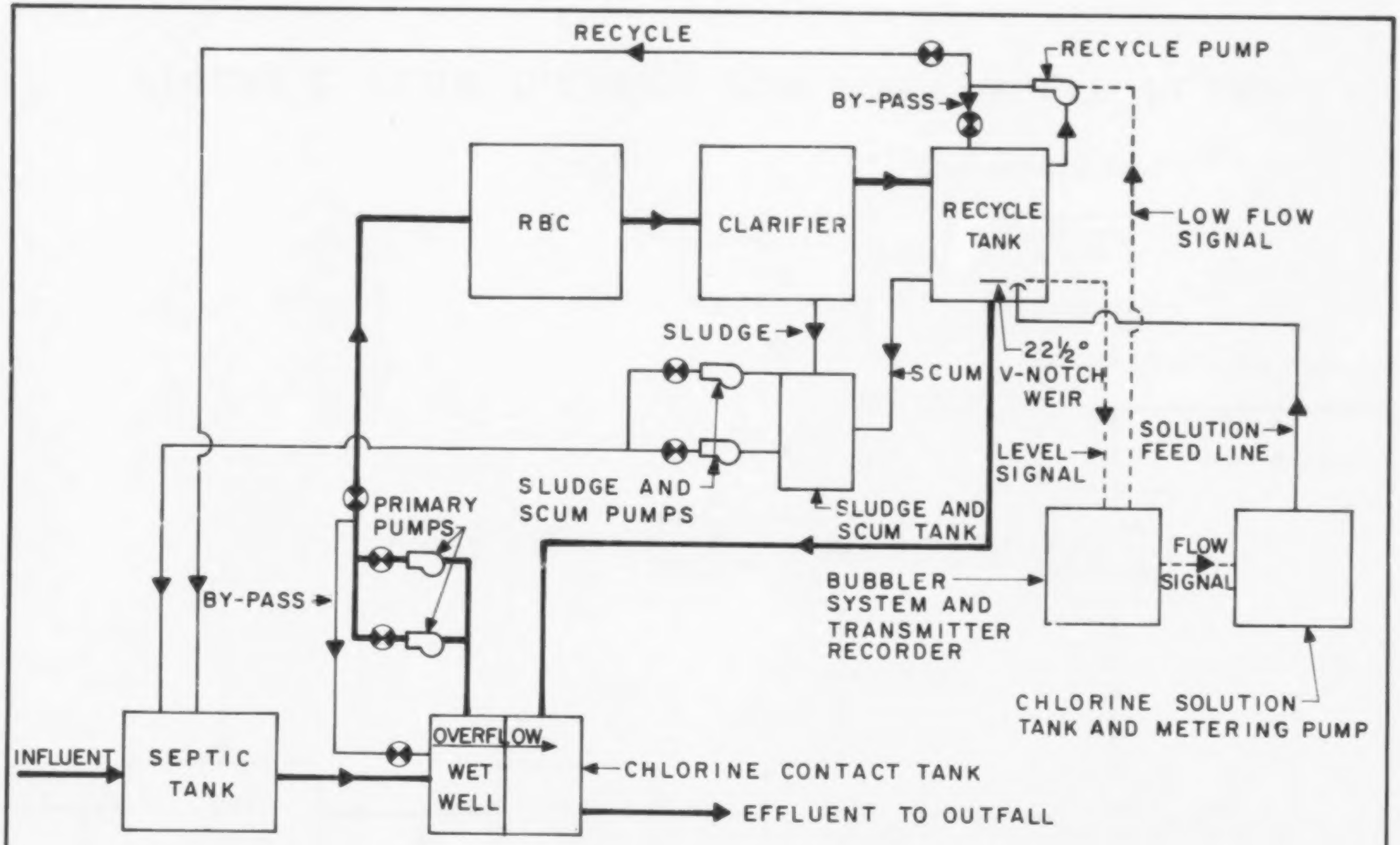
RECYCLE
RECYCLE PUMP
BY-PASS
RBC
CLARIFIER
RECYCLE TANK
LOW FLOW SIGNAL
SLUDGE
SCUM
22½° V-NOTCH WEIR
SOLUTION FEED LINE
LEVEL SIGNAL
SLUDGE AND SCUM PUMPS
SLUDGE AND SCUM TANK
PRIMARY PUMPS
BY-PASS
FLOW SIGNAL
BUBBLER SYSTEM AND TRANSMITTER RECORDER
CHLORINE SOLUTION TANK AND METERING PUMP
INFLUENT
SEPTIC TANK
OVERFLOW
WET WELL
CHLORINE CONTACT TANK
EFFLUENT TO OUTFALL

FIGURE 1 FLOW DIAGRAM FOR HEADQUARTERS PLANT

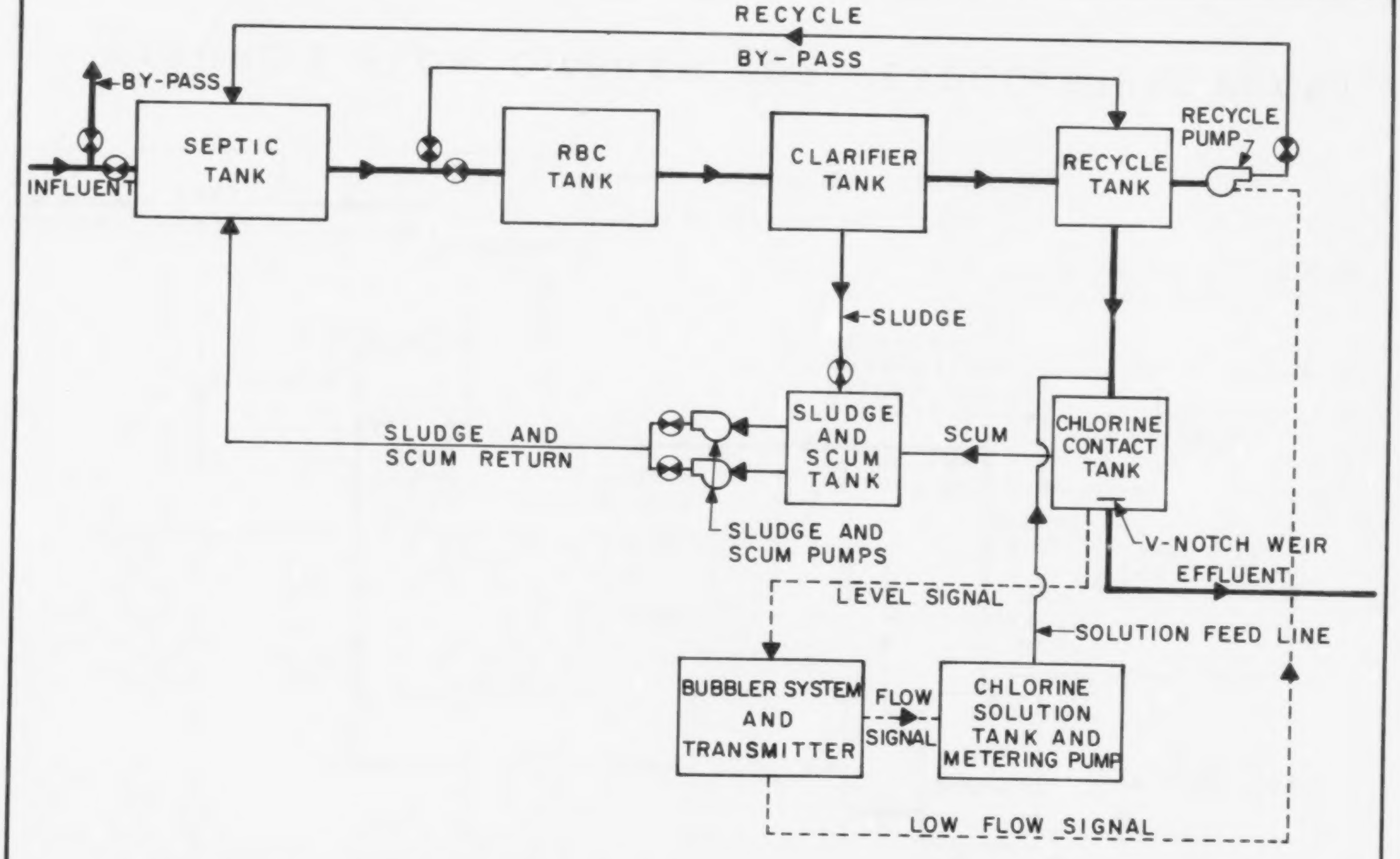
RECYCLE
BY-PASS
BY-PASS
INFLUENT
SEPTIC TANK
RBC TANK
CLARIFIER TANK
RECYCLE TANK
RECYCLE PUMP
SLUDGE
SLUDGE AND SCUM TANK
SCUM
CHLORINE CONTACT TANK
SLUDGE AND SCUM RETURN
SLUDGE AND SCUM PUMPS
V-NOTCH WEIR
EFFLUENT
LEVEL SIGNAL
SOLUTION FEED LINE
BUBBLER SYSTEM AND TRANSMITTER
FLOW SIGNAL
CHLORINE SOLUTION TANK AND METERING PUMP
LOW FLOW SIGNAL

FIGURE 2 FLOW DIAGRAM FOR POINT WOLFE PLANT

THE GLADSTONE, MICHIGAN EXPERIENCE
PERFORMANCE OF A 1.0 MGD RBC PLANT IN A NORTHERN MICHIGAN COMMUNITY

by

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INTRODUCTION

In March of 1974, the first federally funded municipal wastewater treatment plant utilizing Rotating Biological Contactors (RBC) was placed in operation in Gladstone, Michigan. The plant was designed by Williams & Works of Grand Rapids, Michigan to replace an existing primary treatment plant at the same site. Because, in many respects, this plant was the first of its kind in the United States, the performance of the system in a moderately severe northern climate has received a relatively high degree of scrutiny by engineers and manufacturers engaged in the design of other RBC installations. The design considerations for this plant, operating data for four years of operation, and some general observations about RBC installations based on the Gladstone experience will be presented in this paper.

DESIGN CONSIDERATIONS

Gladstone, Michigan is a predominantly residential community of about 5,000 people located in the southern part of Michigan's Upper Peninsula, on the shore of Lake Michigan, approximately 10 miles north of Escanaba. The climate in the area is characterized by cool, dry summers and cold, snowy winters. The mean annual temperature is 42°F (5.5°C). Gladstone derives its municipal water supply from Lake Michigan and wastewater temperatures are frequently in the mid 40's (°F) during winter months.

The 1.0 mgd (3,785 cu m/day) RBC plant was designed to provide secondary treatment and phosphorus removal on the site of an existing primary treatment plant. A system incorporating housed rotating biological contactors was selected on the basis of aesthetics, operational advantages, and data from successful pilot plant testing.

The general design parameters were as follows:

TABLE I
GENERAL DESIGN PARAMETERS
Gladstone, Michigan

Population	10,000
BOD_5	1,670 lb/day (758 kg/day)
TSS	2,000 lb/day (908 kg/day)
Average Daily Flow	1.0 mgd (3,785 cu m/day)
Peak Flow	2.88 mgd (10,000 cu m/day)
Effluent Quality Limitations:	
BOD_5	30 mg/l (30-day average)
Total Suspended Solids (TSS)	30 mg/l (30-day average)
Total Phosphorus (TP)	20% of influent TP

General pilot plant tests of the RBC process were conducted at the University of Michigan in Ann Arbor in 1968-1969 under the supervision of Professor J.A. Borchardt. The pilot plant consisted of three stages in series, each with fifty, 4-foot (1.22 m) diameter, 0.5-inch (1.27 cm) thick was operated for approximately one year using raw wastewater and primary effluent from the Ann Arbor municipal treatment plant. Some results of the pilot plant tests are summarized in Table II. In general, this pilot plant experience offered the following design considerations:

- Primary treatment is necessary to prevent accumulations of debris on RBC shafts and to increase overall BOD_5 removals.
- The RBC shafts must be rotated at a speed sufficient to entrain sloughings and mixed liquor solids. At low speeds, solids accumulations within the RBC tankage caused the depletion of dissolved oxygen and created odor nuisances.
- An average of 80 to 93 percent overall BOD5 removals were consistently achieved at hydraulic loadings of 1 to 4 gpd/sq.ft. and at temperatures ranging from 48^0 to 63^0 F.
- At similar hydraulic loadings, BOD5 reductions through the RBC process dropped about 1% for each 1^0 C drop in wastewater temperature.

A schematic flow diagram of the treatment facility is shown in Figure 1. The design parameters of the unit processes and appurtenant equipment are summarized in Table III. Raw wastewater is screened and pumped to two grit chambers, then comminuted and settled in a single, rectangular, primary clarifier salvaged from the old plant. Primary clarifier effluent is split to two parallel RBC paths, each with three shafts and six stages. Mixed liquor within the RBC tanks flow through 2-foot (0.61 m) diameter submerged openings to subsequent stages. RBC effluent is dosed with liquid alum for phosphorus precipitation prior to pumping to twin rectangular clarifiers. Anionic liquid polymer is added between the pumps and the secondary clarifiers. Secondary effluent is chlorinated prior to discharge to Lake Michigan.

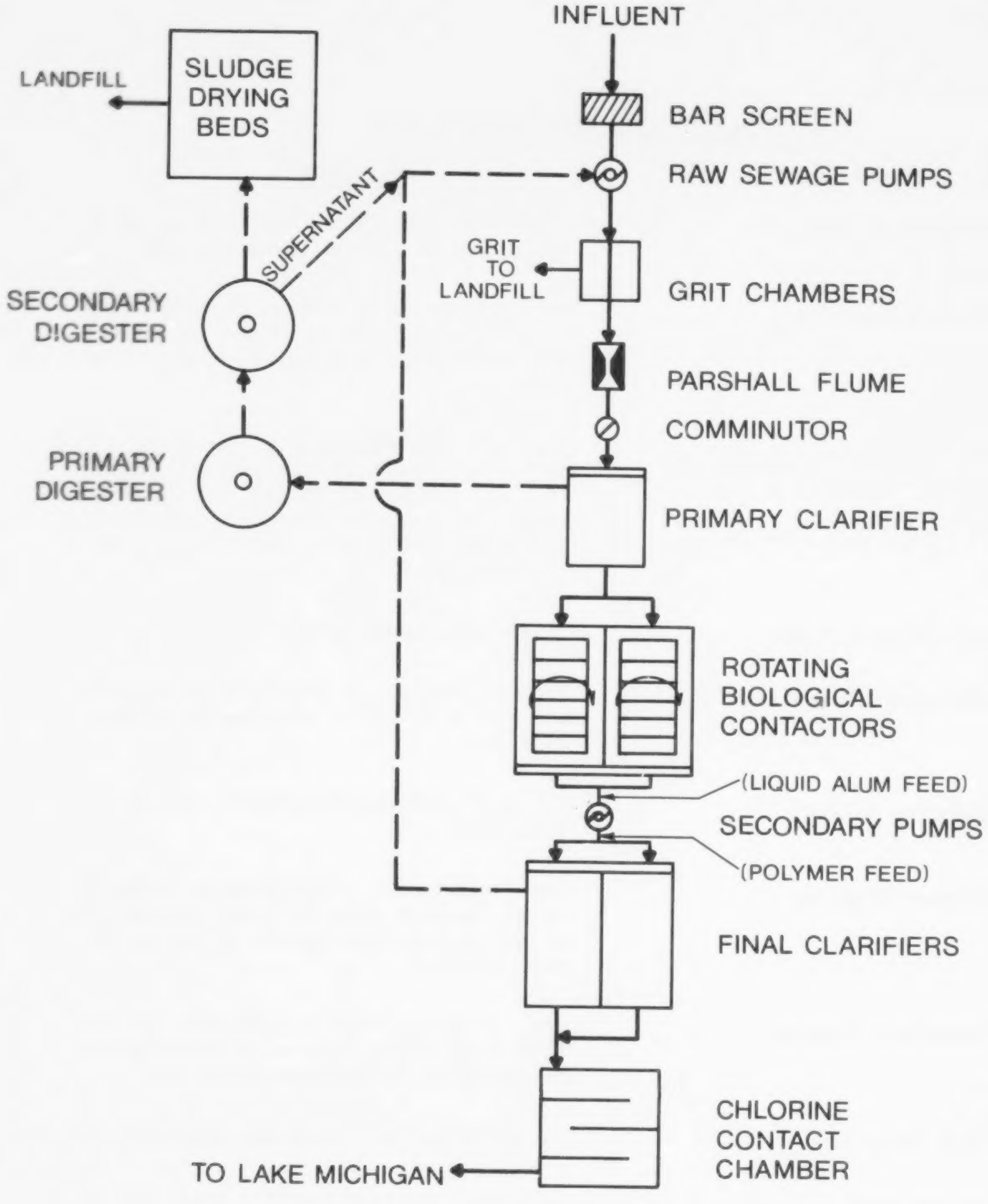
Grit is manually cleaned from the grit chambers and disposed of at a sanitary landfill. Settled primary and secondary sludges are pumped to a primary anaerobic digester that is heated and equipped with gas recirculation and mixing apparatus. Settled, digested sludge from the secondary digester is withdrawn by tank truck to sludge drying beds off-site. Dried sludge is disposed of in a sanitary landfill. Supernatant from the secondary digester is drained by gravity to the raw wastewater wet well.

TABLE II

PILOT PLANT BOD_5 REMOVAL SUMMARY DATA

Test Period	Hydraulic Loading gpd/sq.ft.	Estimated Detention (minutes)	Waste-water Temp. (^{o}F.)	BOD_5 - mg/l					Overall % Removal
				Raw	Prim. Effl.	1st Stage	2nd Stage	3rd Stage	
1/21-2/10/69	4.0 (1,500 gpd)	24	53-59 (56)	96-234 (159)	49-139 (100)	36-76 (59)	18-56 (39)	18-45 (31.9)	73.5-83.5 (80.0)
4/3-4/18/69	2.65 (10,000 gpd)	36	53-59 (56)	103-274 (156)	64-110 (85)	25-54 (42)	12-28 (19)	7-24 (13)	84.5-97.5 (91.6)
5/12-5/21/69	1.32 (5,000 gpd)	72	61-63 (62)	95-144 (116)	44-98 (62)	11-25 (19.2)	8-16 (11.5)	6-13 (8.6)	86.4-95.7 (92.6)
11/17-12/12/69	2.0 (7,500 gpd)	48	48-52 (50)	114-202 (162)	54-198 (111)	37-100 (52.5)	15-65 (32)	11-42 (21.4)	72.2-94.1 (86.8)

Average values are noted in parentheses.


INFLUENT
BAR SCREEN
RAW SEWAGE PUMPS
GRIT TO LANDFILL
GRIT CHAMBERS
PARSHALL FLUME
COMMINUTOR
PRIMARY CLARIFIER
ROTATING BIOLOGICAL CONTACTORS
(LIQUID ALUM FEED)
SECONDARY PUMPS
(POLYMER FEED)
FINAL CLARIFIERS
CHLORINE CONTACT CHAMBER
TO LAKE MICHIGAN
LANDFILL
SLUDGE DRYING BEDS
SUPERNATANT
SECONDARY DIGESTER
PRIMARY DIGESTER

GLADSTONE, MICHIGAN
WASTEWATER TREATMENT SYSTEM
Figure 1

TABLE III

UNIT PROCESS AND EQUIPMENT DATA

Equipment	Design Data
Raw Sewage Pumps	3 @ 1,400 gpm (88.3 l/sec), 2 variable speed, and 1 constant speed lag pump
Primary Clarifier	One rectangular 70,000 gallon (265 cu m) capacity with 1.5 hour detention and 895 gpd/sq ft (36.5 cu m/day/sq m) overflow rate
RBC Units	Bio-Surf by Autotrol. Two parallel paths with three shafts and six stages in each, 515,500 sq ft (47,890 sq m) of media, 90 minutes retention and 1.94 gpd/sq ft (0.097 cu m/day/sq m) hydraulic loading rate
RBC Effluent Pumps	Same as raw sewage pumps
Secondary Clarifiers	Two rectangular, 2.75 hour detention and 620 gpd/sq ft (25.3 cu m/day/sq m)overflow rate
Chlorine Contact	Two baffled tanks with retention time of 35 minutes
Primary Digester	One fixed cover, 100,000 gallon (378.5 cu m) capacity with 15.5 days detention at design loading, heated, mixed by gas recirculation
Secondary Digester	One floating cover, 100,000 gallon (378.5 cu m) capacity with supernatant discharge to raw wastewater wet well
Tank Truck	One 2,500 gallon (9.4 cu m) capacity
Drying Beds	Adjoining sanitary landfill site, 600' x 200' (183 m x 61 m) without underdrains
Liquid Alum Storage	Two tanks, each 2,600 gallon (9.8 cu m) capacity

INITIAL OPERATING EXPERIENCE

The operation of the plant was begun on March 1, 1974. At the same time, an 18-month testing and shakedown program was begun to monitor the performance of the plant and to make operational improvements where necessary. During this test period, wastewater flows averaged 0.755 mgd (2,875 cu m/day) and were influenced occasionally by sewer infiltration and inflows. At start-up, wastewater temperatures averaged 45°F (7.2°C), then gradually increased to the low 60's (°F) by mid-summer.

Influent BOD_5 concentrations ranged from 129 to 219 mg/l and influent total suspended solids concentrations ranged from 101 mg/l to 168 mg/l. The average influent BOD_5 and total suspended solids concentrations for the test period were 164 mg/l and 132 mg/l, respectively.

The results of this initial 18-month test period are summarized as follows:

- Because the influent wastewater temperature was quite cold at startup, it took 18 days before heavy sloughings of biomass were observed and several months to achieve a steady state operation.
- During the last 12 months of the testing period (September 1974 through September 1975) with the addition of 70 mg/l of alum and 0.8 mg/l of polymer, effluent BOD_5's were generally less than 10 mg/l; effluent suspended solids averaged 15 mg/l; and effluent total phosphorus concentrations were less than 1.6 mg/l.
- In January 1975, design flow conditions were simulated for two weeks by shutting down one of the parallel RBC paths and one of the final clarifiers. The effluent BOD_5 for this two-week test period averaged 19 mg/l with an average wastewater temperature of 47°F (8.3°C).
- Because of low hydraulic loadings (1.0 to 2.1 gpd/sq.ft.), nitrification occurred; 50% to 60% reductions in ammonia nitrogen occurred during winter months and 85% to 95% reductions occurred during summer months.
- After one of the RBC paths was restarted in February 1975, after being shut down for full scale tests, nitrifiers did not reestablish dominant cultures on latter stages until June, when wastewater temperatures approached 60° F (15.6°C).

- Excellent BOD5, suspended solids and ammonia nitrogen removals permitted reductions in effluent chlorine doses, from 6 - 7 mg/l to 2 - 3 mg/l.
- Recycling of secondary sludge through the primary clarifier was required to thicken the combined sludges from 2% - 3% solids to 4% - 5% solids content.
- Slug releases of digester supernatant to the head end of the plant were detrimental to overall nitrification results. Controlled release of digester supernatant is considered essential to achieve consistent ammonia nitrogen reductions.
- No significant differences were noticed in the digestibility of sludges with or without alum addition. Phosphorus release, volatile solids reduction, or gas production were not significantly affected by alum addition. The average total sludge production was 0.83-lb of total solids per pound of BOD_5 removed, including chemical sludge production.
- Measured power usage by RBC shafts ranged from 5.6 hp to 3.5 hp. The average running horsepower per shaft was 4.2 hp.

PLANT PERFORMANCE (1976-1979)

The overall performance of the plant, subsequent to the 18-month test program, is summarized in Table IV. The data covers the period from January 1976 through December 1979. Table V shows a comparison of plant performance under summer and winter conditions. Data for these tables was obtained from monthly operating reports prepared by the plant operator.

This data shows that the plant is consistently removing over 90% of influent BOD_5 even under winter conditions and at hydraulic loading rates of between 1.0 and 1.4 gpd/sq.ft. Wastewater strength is affected by increased infiltration and inflow during warm weather months. During winter months, effluent BOD_5 concentrations increase as wastewater temperatures decrease to lows of 45^oF. Effluent dissolved oxygen concentrations are consistently above 7 mg/l.

TABLE IV

PERFORMANCE SUMMARY

Parameter	1976	1977	1978	1979
	- - - - - - Annual	Averages - - - - - -		
Flow (mgd)	.635	.725	.612	.712
Hyd. Loading (gpd/sq.ft.)	1.23	1.41	1.19	1.38
BOD_5 (mg/l)				
- Influent	182	143	155	129
- Effluent	6	7	7	12
- % Removed	97%	95%	95%	91%
TSS (mg/l)				
- Influent	135	122	139	118
- Effluent	15	16	17	16
VSS (mg/l)				
- Influent	111	95	112	92
- Effluent	8	7	8	9
Total Phosphorus (mg/l)				
- Influent	7.8	6.3	4.7	3.5
- Effluent	1.4	1.1	0.9	0.9
- % Removed	82%	83%	81%	74%
Ammonia-N (mg/l)				
- Influent		15.5	17.4	15.7
- Effluent		3.5	2.1	2.7
- % Removed		77%	88%	83%
Effluent Dissolved Oxygen (mg/l)	7.3	7.6	7.6	8.2
Raw Sludge Pumped (lb VSS/day)	576	544	512	544
Digester Gas Prod. (cu.ft./day)	4,307	3,981	3,954	3,718
Chemicals Used (lb/day)				
- Chlorine	13	19	12	12
- Alum	435	485	365	262
- Polymer	1.5	1.7	1.5	1.4

Effluent total suspended solids concentrations remained fairly consistent throughout the period at an average of about 16 mg/l. The hydraulic profile of the plant is such that secondary solids are pumped twice prior to removal with settled primary solids. The addition of polymer assists secondary sedimentation, and lower suspended solids concentrations could probably be obtained in the final effluent if flocculents were not disturbed by centrifugal pumps.

More than 80% of influent total phosphorus concentrations are removed by the addition of alum (hydrated aluminum sulfate). Plant operators attempt to maintain a liquid alum dosage rate sufficient to provide 1.5 moles of aluminum to 1.0 mole of influent total phosphorus.

The plant was not designed to accomplish nitrification. However, because it is currently operated at hydraulic loadings of less than 2 gpd/sq.ft., nitrification does occur. Ammonia analyses on weekly grab samples show that average influent ammonia concentrations of 16 mg/l to 18 mg/l are reduced to 2 mg/l to 4 mg/l through the plant. As might be expected, less nitrification occurs during winter months.

In 1979, an average of 720 pounds per day of sludge was pumped to the primary digester with an average solids content of 3.7% by weight. The volatile content of the sludge averaged 74% by weight. Total sludge production, including chemical sludges, averaged 1.04 pounds per pound of BOD_5 removed, or approximately 1,000 pounds per million gallons treated. Approximately 262 pounds per day of alum was added to remove an average of 2.6 mg/l of total phosphorus. Stoichiometrically, an average of about 90 pounds per day of chemical sludge was produced in 1979.

OPERATING COSTS

Energy and Chemical Costs

The energy and chemical uses and costs associated with the plant operation for 1978 are summarized in Table IV. The costs shown per million gallons reflect actual prices Gladstone paid for the respective commodities in 1978.

TABLE V

COMPARATIVE PERFORMANCE SUMMARY

Winter vs Summer Operation

	Average Winter Values 1/	Average Summer Values 2/
Flow (MGD)	0.530	0.726
Hydraulic Loading (gpd/sq.ft.)	1.03	1.41
Waste Temp (°F)	46	64
BOD_5 (mg/l)		
- Influent	181	128
- Effluent	10	9
- % Removed	94	93
TSS (mg/l)		
- Influent	140	130
- Effluent	16	16
Total Phosphorus (mg/l)		
- Influent	5.8	4.7
- Effluent	0.8	1.0
- % Removed	86	79
Ammonia-N (mg/l)		
- Influent	20.0	15.1
- Effluent	4.9	1.0
- % Removed	76	93
Effluent Dissolved Oxygen (mg/l)	8.1	7.3

1/ From January, February, March, 1977 through 1979 reported data.

2/ From July, August, September, 1977 through 1979 reported data.

TABLE VI

1978 ENERGY AND CHEMICAL USE

Gladstone, Michigan

	Electricity KWH/MG	Fuel 1000 BTU/MG	Chemicals Lbs/MG	Total Cost $/MG
Secondary Treatment				
Raw Sewage Pumping	263	-	-	$ 11.12
RBC System	1,105	-	-	46.74
Chlorination	-	-	20	3.49
Phosphorus Removal				
Alum	-	-	596	27.00
Polymer	-	-	2.4	4.47
Sludge Treatment				
Anaerobic Digesters	40	3,173	-	8.26
Truck	-	105	-	0.52
Drying Beds	-	N	-	N
Miscellaneous				
Building Heat	-	1,923	-	4.00
TOTAL SYSTEM	1,408	5,201	618	$105.60

Average Daily Flow for Year was 0.612 MGD

N = Negligible

Electricity consumption is of primary interest and the table shows electrical power consumption in terms of raw sewage pumping, secondary (RBC) treatment, and anaerobic digestion. The power consumption for the RBC units actually includes power consumption by clarifier drives, intermediate pumps, lights, ventilating equipment, etc. and computes to an average of approximately 38 running horsepower at any given time. Of this, approximately 80 percent (30 hp), including power draw by the RBC units, could be considered independent of actual flows. Based on measured power consumptions by the six RBC drive units in 1975, the RBC system draws an average of 25-26 horsepower at any given time.

To gain more perspective on energy consumption at Gladstone, as related to consumptions by other types of small wastewater treatment plants, Table VII displays similar breakdowns for six treatment plants in Michigan. Because of wide variations in commodity costs from community to community, the costs shown in Table VII have been adjusted to a common base as follows:

Electricity	$0.042 per KWH
Fuel	
Natural Gas	$0.210 per 100 cu. ft.
Gasoline	$0.700 per gallon
Chemicals	
Chlorine - Effluent	$0.120 per pound
- Alma Purifax	$0.097 per pound
Lime	$0.036 per pound
Alum	$0.050 per pound
Ferric Chloride	$0.073 per pound
Polymer	$1.880 per pound

It is noteworthy that the net energy usage by the Alma extended aeration system and the Gladstone RBC system are about the same. However, the Gladstone plant was currently only receiving about 60 percent of its design flow, whereas the Alma system was at about 94 percent of capacity. Net energy usage by the two systems at design flow conditions, with appropriate assumptions about variable power usage, is forecast as follows:

TABLE VII

COMPARISON OF 1978 ENERGY COST FOR DIFFERENT TYPES OF TREATMENT*

	Raw Sewage Pumping Cost $/MG	Secondary Treatment Cost $/MG	Phosphorus Removal Cost $/MG	AWT Cost $/MG	Sludge Treatment Cost $/MG	Total Cost $/MG
Vermontville (0.067 MGD) - Oxidation lagoons - Flood Irrigation	34.06	0	0	0	0	34.06
Middleville (0.213 MGD) - Oxidation Lagoons - Spray Irrigation	36.79	0	0	39.79	0	76.58
Jackson Co.-Leoni (0.862 MGD) - Aerated Lagoons - Spray Irrigation	16.72	26.17	0	20.26	0	63.15
Ludington (2.01 MGD) - Aerated Lagoons - Chemical Precipitation	23.20	29.36	65.06	NA	+	117.44
Alma (2.35 MGD) - Extended Aeration - Polishing Pond - Purifax	8.95	54.44	17.47	0	12.02	92.88
Gladstone (0.612 MGD) - RBC System - Anaerobic Digestion	11.04	52.87	34.27	NA	8.86	107.04

NA - Not Applicable to this system.

* The costs for all treatment systems are adjusted for the same unit costs.

+ Lime addition for phosphorus removal aids sludge treatments.

	1978 Flow % of Design	1978 Electrical Use, KWH/MG	Adjusted Electrical Use at Design Flow, KWH/MG
Alma	94	1,209	1,150
Gladstone	61	1,105	700-800

Thus, electrical power use for secondary treatment at Gladstone should be 30 percent to 35 percent less than that at Alma as the design flows for each plant are approached. Comparison of the two plants must be tempered by an appreciation for higher degrees of treatment achieved at Alma.

Overall Operation and Maintenance Costs

Table VIII is a conceptual operation and maintenance budget for a system such as Gladstone's for fiscal year 1980. The Gladstone plant is staffed by one superintendent/operator and four shift operators. The plant is manned 16 hours per day, five days per week; eight hours per day on weekends and holidays. Approximately 30 percent of the labor budget at Gladstone is allocated to the water sypply system; however, the conceptual budget assumes total assignment of labor to the wastewater treatment system. Labor costs for the five-man staff are estimates for 1980, but are considered reasonable for current average operating labor in this region of the country. Energy and chemical costs reflect Gladstone's 1978 costs, increased by 20 percent. The equipment replacement line item is in deference to current EPA user charge requirements for funds to replace key process equipment items over the service life of a facility. This budget suggests a treatment cost of approximately $600 per million gallons of wastewater treated.

SUMMARY

In summary, the Gladstone experience has been a good one. Excellent treatment results are being obtained; plant effluent quality is well within design limits. Consistent nitrification at loadings below 2 gpd/sq.ft. is being achieved as an added bonus toward high quality treatment.

The RBC system has been very consistent in achieving wastewater treatment and in permitting stable solids handling procedures at Gladstone. Few, if

TABLE VIII

CONCEPTUAL OPERATION AND MAINTENANCE BUDGET

FY 1980

Labor/Benefits	$ 96,000
Power/Fuels/Chemicals	33,000
Routine Plant Maintenance	3,000
Parts Inventory/Supplies/Miscellaneous	4,000
Equipment Rental	3,000
Equipment Replacement Fund	15,000
TOTAL	$154,000
Cost per million gallons treated	$595/MG[1/]

[1/] Based on treating 260 MG per year.

any, problems have occurred with the day-to-day operation of this RBC secondary treatment system. The process has been economical and has required a minimum of operator attention over the past four years.

ACKNOWLEDGEMENTS

The authors wish to thank the City management of Gladstone, Michigan and Mr. Willard Morley, Plant Superintendent, for making plant operation data available for this paper.

REFERENCES

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PART X: NITRIFICATION AND DENITRIFICATION

The Effect of Organic Loading on Nitrification in RBC Wastewater Treatment Processes

by

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INTRODUCTION

Though organic substances, themselves, are not harmful to nitrifying bacteria, the nitrification in the biofilm attached to the rotating disc may be influenced by them. Under high organic loading, the number of nitrifying bacteria in the biofilm will be decreased, not because of substrates, dissolved oxygen, or essential nutrients competition, but because of the difference of growth rate between the nitrifying bacteria and the aerobic heterotrophs. The aerobic heterotrophs will grow very rapidly under high organic loading, and the biofilm on the disc will thicken and eventually be sloughed off by gravity and shearing stresses acting on the rotating biofilm. As a result of frequent sloughing, the nitrifying bacteria may be washed out from the attached biofilm.

One prominent difference between the ordinary conventional trickling filter and the rotating biological contactor (RBC) is the longitudinal mixing characteristics of their flow patterns. In the trickling filter, the flow pattern is thought to be that of plug flow. But in the RBC, the flow pattern may be closer to that of complete mixing. Because of this difference, nitrification in the RBC may be more sensitive to effects of variations of organic loading.

In this paper, some results of experiments are presented to discuss to what extent organic loading may affect nitrification, how to maintain the stable nitrification in the RBC, and how to proceed with denitrification in it.

EXPERIMENTAL EQUIPMENT AND METHOD

Three different types of continuous flow laboratory scale RBC units were used. One was a four-stage RBC unit and the others were single-stage RBC units. Schematic flow diagrams are shown in Figs.1 and 2, respectively. Dimensions and other specifications are given in Table 1.

Six series of experiments were carried out. Classifications of experiments and the composition of influents used are briefly summarized in Table 2.

Analytical methods for determining water quality parameters used in these experiments and techniques of bacterial enumeration are summarized in Table 3. The amount of biomass attached to the discs was weighed directly after half a minute of dewatering. The thickness of biofilm near the rim of a disc was measured at several points by slide caliper. Biomass sloughed off from discs was retrieved with a fine mesh nylon net, dried, and weighed to evaluate the rate of biomass sloughing in the RBC.

Table 1. Dimensions of the RBC units

RBC unit classification	Type(A)	Type(B)	Type(C)
Number of stages	4	1	1
Number of discs	10 x 4	10	5
Disc Diameter	25 cm	25 cm	30 cm
Material	PVC	PVC	PVC
Interval between discs	1 cm	1 cm	5 cm
Contactor volume	28.8 litres (7.2 x 4)	7.2 litres	10.7 litres
Fraction of area submerged	0.45	0.45	0.49
Rotational speed	15 rpm	15 rpm	2-10 rpm

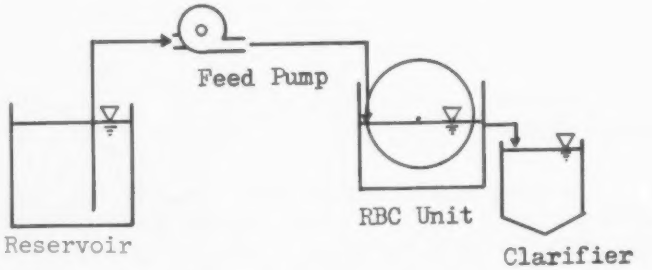

Fig. 2. Schematic flow diagram of the Type(B) or Type(C) contactor

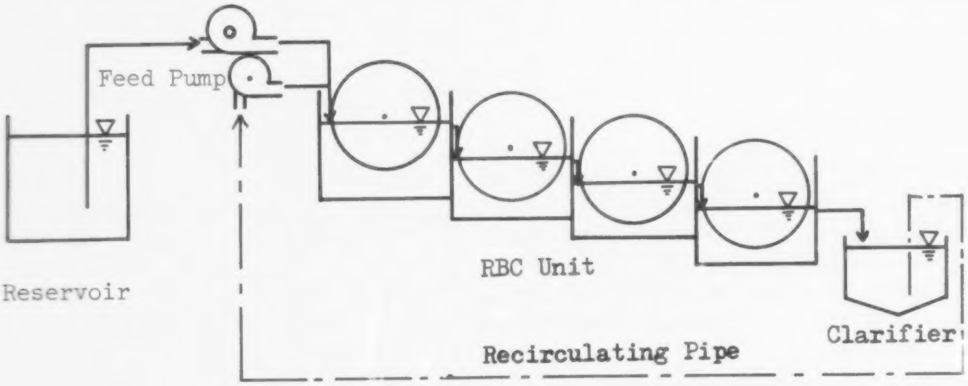

Fig. 1. Schematic flow diagram of the Type(A) contactor

Table 2. Classifications of experiments

Classifications of experiments	RBC unit type	Composition of influent sources
(1) Experiment to estimate the growth rate of nitrifying bacteria	Type(B)	$(NH_4)_2SO_4$, Na_2CO_3 nutrient salts soluted in tap water
(2) Experiment to calculate the biomass sloughing rate	Type(C)	synthetic dry milk* soluted in tap water
(3) Experiment to estimate critical organic loading for nitrification	Type(A)	$(NH_4)_2SO_4$, Na_2CO_3 synthetic dry milk soluted in tap water
(4) Experiment to count numbers of nitrifying bacteria	Type(A)	$(NH_4)_2SO_4$, Na_2CO_3 synthetic dry milk soluted in tap water
(5) Experiment on denitrification using a recirculating system	Type(A) with recirculation	$(NH_4)_2SO_4$, Na_2CO_3 synthetic dry milk soluted in tap water
(6) Experiment to proceed with both nitrification and denitrification	Type(B)	NH_4Cl, $NaHCO_3$ nutrient salts glucose or methanol soluted in tap water

* synthetic dry milk composition: C 50 %, N 2 %, P 0.4 %
several vitamins and minerals are included

Table 3 Analytical methods

Measurement	Analytical method
Organic nitrogen	Kjeldahl method
Ammonium nitrogen	Ammonium electrode method
Nitrate nitrogen	Cadmium column reduction method
Nitrite nitrogen	GR method
COD(Cr)	Standard Methods
TOC	Standard Methods
Dissolved oxygen	Azide modification of Winkler method
pH	Glass electrode method
Viable count of nitrifying bacteria	MPN technique
Viable count of aerobic heterotrophic bacteria	Spread plate technique with Sakurai's agar medium

EXPERIMENTAL RESULTS

Observed growth rate of nitrifying bacteria on discs

Concerning environmental conditions for nitrification, dissolved oxygen concentration and pH were maintained higher than 5 mg/l and between 6.8 and 7.8, respectively. The water temperature was held between 15°C and 25°C, and the influent ammonium was mostly oxidized to nitrate. The thickness of biofilm, assumed to consist mainly of nitrifying bacteria, was very thin, less than 1 mm, and rather uniform over the surface of a disc. The sloughing rate of the biomass was less than 0.1 gSS/day·m^2 as shown in Fig.3.

From results of viable counts of ammonium oxidizing and nitrite oxidizing bacteria in the biofilm, we could estimate the number of these bacteria at 10^7 to 10^9/m^2 and 10^7 to 10^8/m^2, respectively. The relationship between the ammonium nitrogen removal rate and the ammonium nitrogen loading is shown in Fig.4. The obtained maximum removal rate was 4 gNH$_4$-N/day·m^2. A comparison between the observed specific growth rate of nitrifying bacteria and the observed specific removal rate of ammonium nutrogen is shown in Fig.5. The yield coefficient could be determined to be 0.15 from these results. It was significant that both the maximum specific growth rate of nitrifying bacteria and the maximum specific removal rate

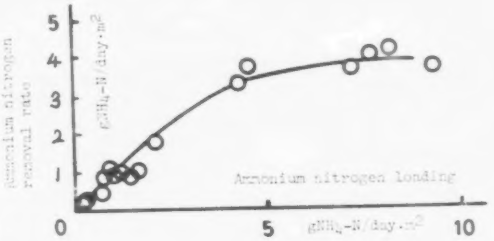

Fig.4. Relationships between ammonium nitrogen removal rate and ammonium nitrogen loading

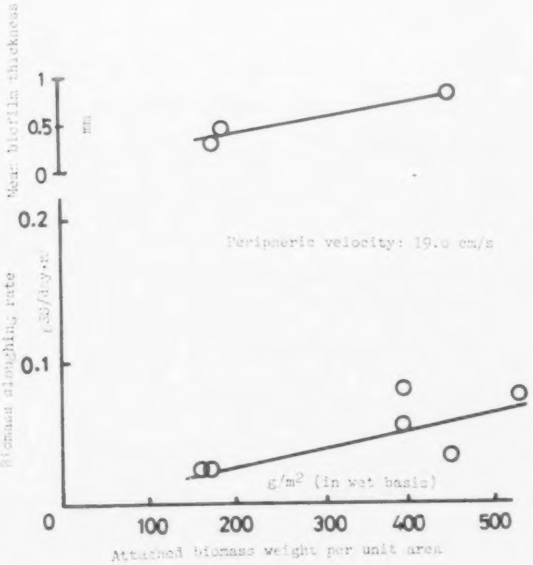

Fig.3. Sloughing rate and mean biofilm thickness

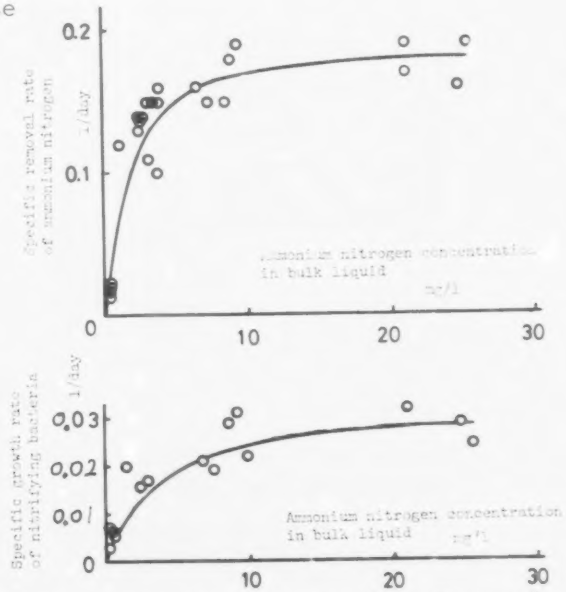

Fig.5. Comparison between the observed specific growth rate of nitrifying bacteria and the observed specific removal rate of ammonium nitrogen

of ammonium nitrogen in the biofilm on the disc were as low as 0.03 and 0.2/day, respectively, and these values were about one tenth of those(1)(2)(3) in dispersed cultures, respectively.

Biomass sloughing rate

In this series of experiments, water temperature and pH were maintained between 18°C and 20°C and between 7.0 and 7.5, respectively. In the case of the rotational speed being 9.7 rpm, the dissolved oxygen concentration was greater than 3 mg/l.

As the organic loading was increased, the thickness of biofilm increased rapidly and the dominant species changed into filamentous organisms. Relationships between the attached biomass weight and the organic loading expressed in COD(Cr), and the sloughing rate of the biomass from biofilms are shown in Fig.6 and 7, respectively. When the COD loading was increased to higher than 20 g/day·m^2, the sloughing rate increased to 2 gSS/day·m^2. Generally speaking, biofilms grew very rapidly but at the same time they were sloughed off very frequently.

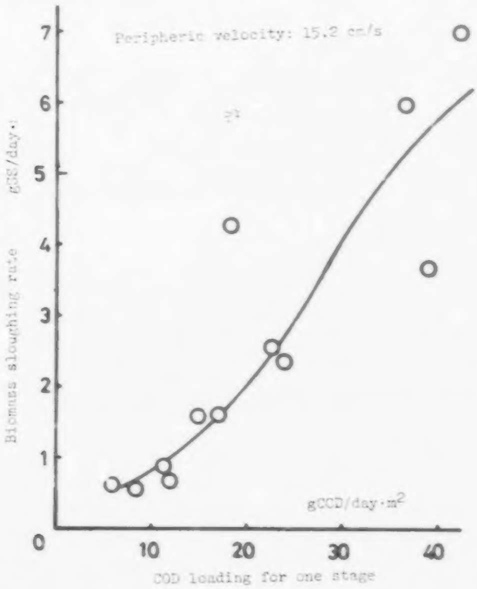

Fig.7. Relationship between biomass sloughing rate and organic loading(COD loading)

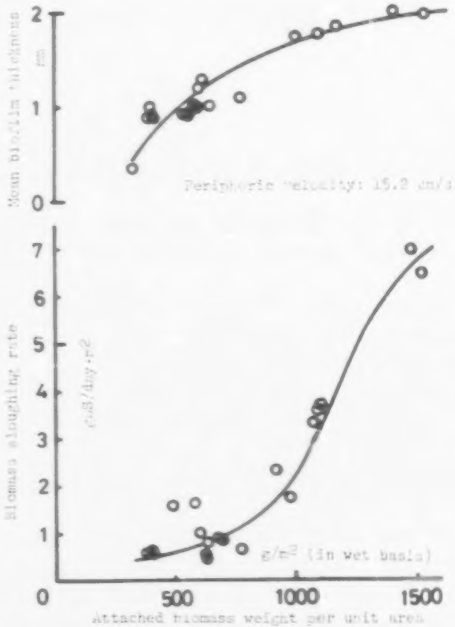

Fig.6. Relationships between biomass sloughing rate,mean biofilm thickness, and attached biomass weight

Regarding some effects of rotational speed, the sloughing rate, the COD removal rate, and the attached biomass weight were measured under three different rotational speeds. In this experiment, both COD loading and hydraulic loading were kept constant at 35 gCOD/day·m^2 and 0.11 m^3/day·m^2, respectively. Results are shown in Fig.8.

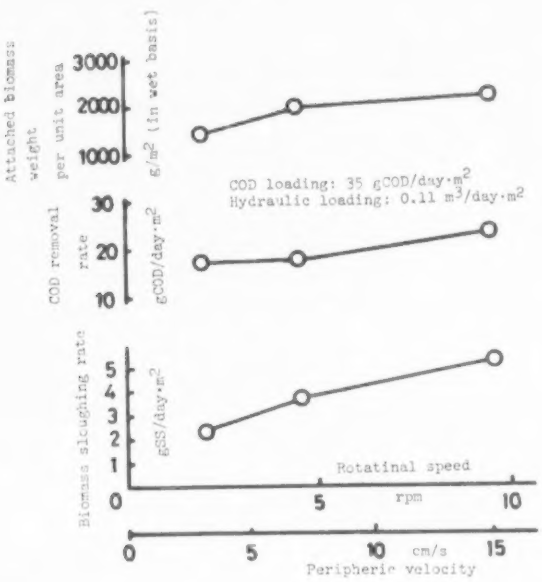

Fig.8. Effects of rotational speed

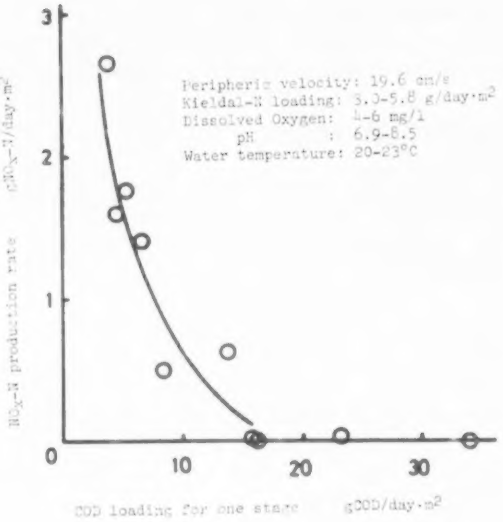

Fig.9. Critical organic loading for nitrification

Critical organic loading for nitrification

A relationship between the production rate of oxidized nitrogen (NO_x-N = NO_2-N + NO_3-N) and COD loading is shown in Fig.9.

From the balance of removed total nitrogen, produced NO_x-N, and nitrogen taken into increased biomass, we could not find any evidence of the occurrence of denitrification. Therefore, the production rate of NO_x-N was nearly equal to the nitrification rate and it was concluded that nitrification could not proceed when COD loading was increased to higher than 20 gCOD/day·m².

Viable counts of nitrifying bacteria

The time dependent variation of viable counts of nitrifying bacteria and aerobic heterotrophic bacteria in biofilms were examined. This was carried out using the second stage unit of a four-stage RBC, because it was necessary that there exist simultaneously a certain amount of nitrifying bacteria and aerobic heterotrophic bacteria as an initial condition. After the organic loading was increased suddenly, the biofilm grew very rapidly and frequent sloughing began. As the amount of biomass on discs increased, viable counts of aerobic heterotrophic bacteria increased, but those of nitrifying bacteria decreased as shown in Fig.10.

Denitrification by the use of a recirculating system

As seen previously, denitrification did not occur in the usual one-through operation. There have been some reports which give attempts to proceed with denitrification by recirculating the effluent to the first stage of the RBC. In the present series of experiments, correlations among the removal efficiency of NO_x-N, the recirculation ratio, and the organic loading were investigated concerning the removal of total nitrogen from the water as a whole system.

A relationship between the NO_x-N removal efficiency and COD loading, which was loaded on the first stage under the constant hydraulic loading of 0.08 m^3/day·m^2, is shown in Fig.11. Although the observed NO_x-N removal efficiency is not identical to the denitrification efficiency in a strict sense, we can see that denitrification proceeded very well under higher organic loading than 20 gCOD/day·m^2. We can expect about 90% NO_x-N removal under the COD loading of 40 gCOD/day·m^2 in the first stage.

The total nitrogen (T-N) removal efficiency through a whole stage was calculated from the balance between concentration of T-N of influent and that of effluent. A comparison of results with and without recirculation is shown in Fig.12.

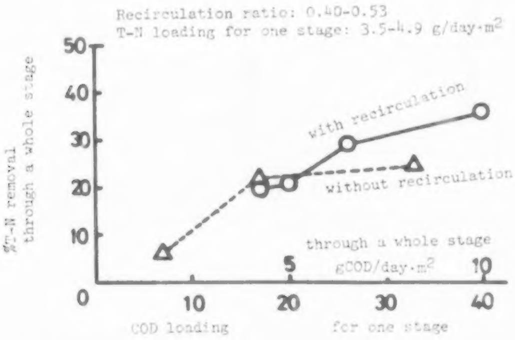

Fig.12. Comparison of T-N removal efficiency through a whole stage with and without recirculation

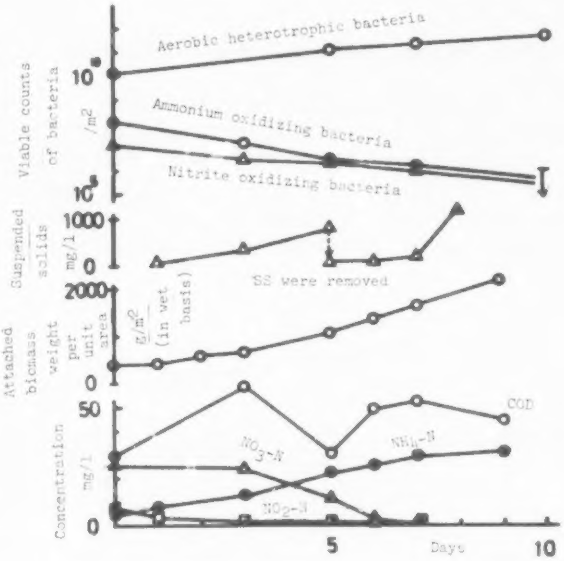

Fig.10. Variation of viable counts of nitrifying bacteria with time under sudden high organic loading

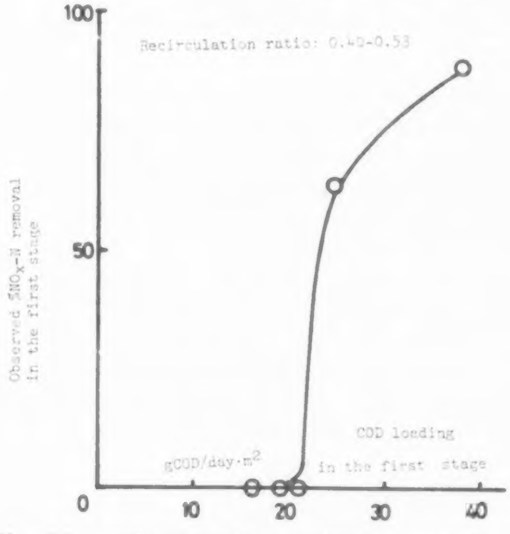

Fig.11. relationship between NO_x-N removal efficiency and COD loading in the first stage

In the case without recirculation, the T-N removed from water might be converted to biomass. Accordingly, we can expect about a 10 per cent increase in T-N removal efficiency by denitrification of the recirculated water in the first stage. Relationships between the recirculation ratio and the NO_x-N removal efficiency in the first stage, T-N removal efficiency through a whole stage, and concentration of ammonium nitrogen (NH_4-N) in effluent water are shown in Fig.13.

Nitrification and denitrification on the same disc

To investigate the possibility of simultaneously proceeding with both reaction of nitrification and denitrification in the biofilm on the same disc, a preliminary experiment was carried out. According to results of this experiment as shown in Fig.14, we could not confirm this possibility when glucose was used as an organic source. However, confirmation was possible, as anticipated, when methanol was used as the organic source.

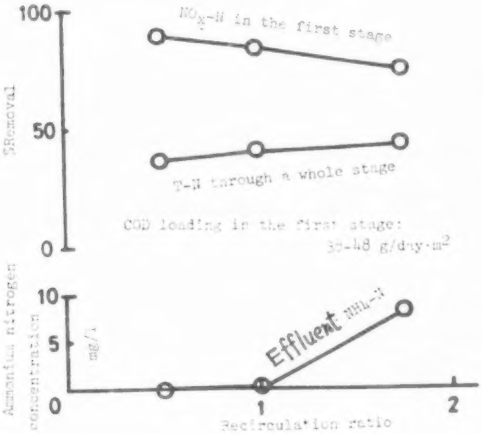

Fig.13. Relationships between recirculation ratio and NO_x-N removal efficiency, T-N removal efficiency, and effluent NH_4-N concentration

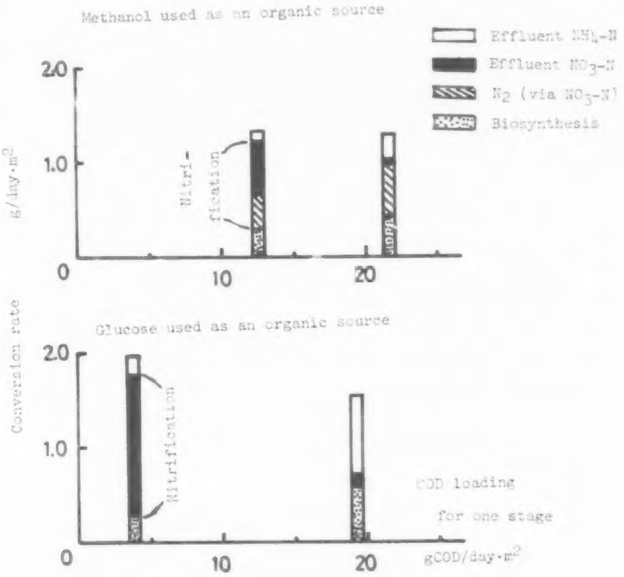

Fig.14. Conversion of influent ammonium nitrogen with methanol and glucose used as an organic source

DISCUSSION

The obtained yield coefficient of nitrifying bacteria in this study is nearly equal to the value of 0.16 which can be calculated from the following empirical equations.(5)

Nitrosomonas

$$55\ NH_4^+ + 76\ O_2 + 5\ CO_2 = C_5H_7O_2N + 54\ NO_2^- + 109\ H^+ + 52\ H_2O$$

Nitrobacter

$$400\ NO_2^- + 195\ O_2$$
$$+ 5\ CO_2 + NH_4^+ + 2\ H_2O = C_5H_7O_2N + 400 NO_3^- + H^+$$

However, the observed maximum specific growth rate of nitrifying bacteria was apparently small. The effective thickness of a biofilm for nitrification was estimated using the mathematical kinetic model as summarized in Tables 4 and 5.

Table 4. Mathematical kinetic model for zero order reaction

	Ammonium nitrogen	Dissolved oxygen
Mass balance equation	$D_b \frac{d^2 C_i}{dz^2} = R$	$D_b^* \frac{d^2 C_i^*}{dz^2} = R^* = kR$
Boundary conditions	$\frac{dC_i}{dz} = 0$ at $z = L_e$, $C_i = C_{io}$ at $z = 0$	$\frac{dC_i^*}{dz}$, $C_i^* = 0$ at $Z = L_e$
Substrate concentration for zero order reaction	$C_i = \frac{\mu_m X}{2D_b Y} z^2 - \frac{\mu_m X L_e}{D_b Y} z + C_{io}$	$C_i^* = \frac{k\mu_m X}{2D_b^* Y} (z^2 - 2L_e z + L_e^2)$
Effective biofilm depth (L_e)	$L_e = \frac{NY}{\mu_m X}$ (1)	$L_e = \frac{N^* Y}{k\mu_m X}$ $= - \frac{D_b^*}{D_w^*} L_1 + \sqrt{(\frac{L_1 D_b^*}{D_w^*})^2 + \frac{2YD_b^* C_o^*}{k\mu_m X}}$ (2)
Remarks	developed and listed in different expressions by Williamson and McCarty(5)	assuming that the removal rate of oxygen is proportional to that of ammonium nitrogen and the dissolved oxygen concentration decreases to 0 mg/l under the oxygen transfer limitation
Nomenclature	C_i, C_i^* = concentration of ammonium nitrogen and oxygen in the biofilm,respectively $[M/L^3]$ R, R^* = reaction rate of ammonium nitrogen and oxygen,respectively $[M/L^3T]$ k = proportional constant D_b, D_b^* = diffusion coefficient within the biofilm for ammonium nitrogen and oxygen, respectively $[L^2/T]$ z = distance into biofilm from the biofilm surface $[L]$ L_e = effective biofilm depth $[L]$ C_{io} = concentration of ammonium nitrogen at the biofilm surface $[M/L^3]$ μ_m = the maximum specific growth rate of nitrifying bacteria $[1/T]$ Y = yield coefficient X = bacterial concentration in the biofilm $[M/L^3]$ N, N^* = surface flux of ammonium nitrogen and oxygen,respectively $[M/T \cdot L^3]$ D_w^* = diffusion coefficient through water for oxygen $[L^2/T]$ L_1 = stagnant liquid layer depth $[L]$ C_o^* = dissolved oxygen concentration in bulk liquid $[M/L^3]$	

Table 5. Effective thickness of the biofilm calculated from Eqs.(1)and(2)

	Eq. (1)	Eq. (2)
Effective biofilm thickness (L_e)	30 - 110 μm	20 - 120 μm
Terminal concentration of ammonium nitrogen (C_i at $z = L_e$) when the concentration of ammonium nitrogen in the bulk liquid is 10 mg/l	5 - 9 mg/l	6 - 8 mg/l
Parameter values used $Y = 0.16$ $k = 4.25$ Area factor = 1.0 $C_o^* = 5 - 8$ mg/l	$D_b = 1.4$ cm^2/day(5) $D_b^*, D_w^* = 2.5$ cm^2/day(5) $\mu_m = 0.2 - 0.4$ /day(2)(4) $X = 30 - 50$ mg/cm^3(6)(7) $L_1 = 50 - 100$ μm(5)(6)(7)	
Diffusion coefficient through water for ammonium nitrogen = 1.4 cm^2/day(5)		

These tables show that the biofilm is under the condition of oxygen transfer limitation, and it can be inferred that the effective thickness was approximately 50 μm to 100 μm, about one tenth of a whole thickness of the biofilm, which explains the small value of the observed specific growth rate. Therefore, it can be assumed that nitrification can proceed only in the very thin surface layer of a biofilm, and that the continuous rapid growth of aerobic heterotrophs and the frequent sloughing of biofilms should eventually cause the extermination of nitrification in the unit. As a numerical value of the critical organic loading which exterminates nitrification, a value of 20 gCOD/day$\cdot$m^2 was obtained for one stage of the RBC used in this study. At this organic loading, the sloughing rate of biofilms reached 2 gSS/day$\cdot$m^2. This rate was far beyond the maximum growth rate of nitrifying bacteria of 0.7 gSS/day$\cdot$m^2 which was calculated from obtained values of the maximum removal rate of ammonium nitrogen and the yield coefficient.

It seems very promising to proceed with denitrification by using organic substances in influent water. However, it appears necessary to maintain a high organic loading, more than 20 gCOD/day$\cdot$m^2 for one stage of the RBC used in this study, in order to proceed with denitrification. For the case of high organic loading, it may be very effective to recirculate the effluent to the first stage of a multistage RBC to remove nitrogen without any other addition of organic substances and special facilities. When methanol only was used as the organic material, the biofilm appeared to be stable enough to maintain the nitrifying bacteria in the surface layer and the denitrifying bacteria in the internal layer because of the rather small growth rate of the aerobic methylotrophs. In this special case, both nitrification and denitrification may proceed simultaneously on the same disc.

SUMMARY

The results obtained above can be summarized as follows:

(1) Nitrification proceeds mainly in the surface layer of the biofilm and nitrifying bacteria in this layer may be washed out easily by frequent sloughing of biofilms under higher organic loading.

(2) As nitrification is rather sensitive to organic loading, it is recommended that multistage RBC be used where each stage is connected by cascade to prevent the ill-effects of intermixing and to maintain stable conditions for nitrification.

(3) Under higher organic loading, the recirculation of the effluent to the first stage unit may proceed with denitrification and will improve the nitrogen removal efficiency.

(4) Both reactions of nitrification and denitrification can be done simultaneously by the use of methanol as the only organic source, but this may not be expected when other ordinary organic substances are employed.

ACKNOWLEDGMENT

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NITRIFICATION ENHANCEMENT THROUGH PH CONTROL WITH ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

Recent emphasis on nitrogen control in receiving bodies of water has required the removal of ammonia nitrogen from the effluent of many wastewater treatment facilities. The desire to utilize wastewater treatment technology compatible with existing facilities, minimize energy requirements, insure ease of operation, and achieve the desired ammonia removal efficiencies has resulted in the current examination of rotating biological contactor (RBC) systems for possible widespread utilization in such tertiary treatment system applications (1-7).

One of the potential problems associated with such fixed film nitrifying systems is the reduced rate of nitrification which occurs when treating low alkalinity-low pH wastewaters (1,2). It is the intent of this paper to discuss a current research effort which will ultimately assist in the development of design criteria for RBC systems for use in nitrifying low alkalinity-low pH wastewaters.

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BACKGROUND

The organisms which convert ammonia to nitrate are mainly a mixed population of chemoautotrophic bacteria which initially convert ammonia to nitrite and then to nitrate. Energy is derived from this process which is used to support other metabolic processes. The Nitrosomonas and Nitrobacter bacteria are the two primary genera responsible for this nitrification process. The respective oxidation reactions are presented below (1). As can be seen from these reactions, the nitrifying process results in the creation of acid which effectively neutralizes the alkalinity in the wastewater. Theoretically, 7.1 mg/ℓ of alkalinity is destroyed for each mg/ℓ of ammonia oxidized. The destruction of alkalinity results in an increasing pH depression as the amount of ammonia oxidized becomes greater. The actual magnitude of the pH depression is somewhat mitigated by the removal of carbonic acid through the stripping of CO_2 from the wastewater surface (1).

Ammonia Oxidation:

$$NH_4^+ + 1.5\ O_2 + 2\ HCO_3^- \xrightarrow{\text{Nitrosomonas}} NO_2^- + 2\ H_2CO_3 + H_2O \quad (1)$$

Nitrite Oxidation:

$$NO_2^- + 0.5\ O_2 \xrightarrow{\text{Nitrobacter}} NO_3^- \quad (2)$$

Overall Reaction:

$$NH_4^+ + 2\ O_2 + 2\ HCO_3^- \longrightarrow NO_3^- + 2\ H_2CO_3 + H_2O \quad (3)$$

The observed level of alkalinity varys significantly among wastewaters. The major influencing factor in the amount of alkalinity present is the drinking water supply which ultimately discharges into the wastewater treatment system. Relatively high natural alkalinities are associated with certain ground water supplies and much lower alkalinities normally are associated with surface supplies. Domestic usage adds approximately 100 mg $CaCO_3$/ℓ to the natural alkalinity of the carriage water (8). Therefore, the amount of alkalinity in a domestic wastewater may range from 100 mg $CaCO_3$/ℓ or less to several hundred mg $CaCO_3$/ℓ. The net effect of such variations in alkalinity is to provide a different buffering capacity for each wastewater treatment system. Domestic wastewaters may have from about 12 to 25 mg/ℓ of ammonia nitrogen. The range of alkalinity destroyed during the nitrification of such ammonia concentrations is 85 mg $CaCO_3$/ℓ to 178 mg $CaCO_3$/ℓ. Obviously the pH depression can be very slight for low ammonia-high alkalinity wastewaters or very significant for high ammonia-low alkalinity wastewaters. The effect of this relationship on nitrification becomes more significant as the amount of alkalinity drops or the level of ammonia increases.

A number of publications have, to varying degrees, addressed the effect of pH on the nitrification process (1,2,3,7,9,10,11,12). The reported data show a wide variety of optimum pH. In general, the nitrifying organisms function in an optimum mode at approximately pH 8.0 to 8.5 The nitrifying performance then decreases with decreasing pH to about pH 5.5 where nitrification essentially ceases. The reported rates of nitrification are relatively ambiguous between pH 5.5 and 8.0. Unfortunately, data gathered to date on RBC wastewater treatment systems largely have failed to establish the difference between long and short term effects. The research efforts on the effect of pH on nitrification with wastewater treatment systems in most cases have been short term studies (10). One of the problems involved in investigating the nitrification process is the slow growth rates of the organisms. This factor has been a major reason for the inability to date to properly establish the long term effect of pH change relative to short term pH changes. The ability of an organism to successfully acclimate to lower pH values has a significant bearing on the degree of biological nitrification which is to be expected for any particular wastewater treatment system. If engineers are to be able to design RBC biological nitrification systems which optimize the rate of nitrification, more definitive research is necessary to clearly establish the relative rates of nitrification for both short term and long term pH changes. Only after the long term nitrification rates have been established can the benefit of pH adjustment be made clear.

The literature also fails to address the efficacy of alternative chemicals and chemical feed schemes for the adjustment of pH in RBC systems in order to achieve optimum nitrification. Such pH adjustments have been attempted for both RBC and activated sludge systems (2,13). Unfortunately, these studies have been inconclusive in establishing the benefit of chemical addition. Studies on RBC systems which have treated high strength ammonia wastewaters have reported success in practicing alkaline addition (14,15). However for those RBC facilities which are designed to nitrify low alkalinity-low pH domestic wastewaters, additional research in this area would assist design engineers in their attempts to provide the greatest rate of nitrification at the least initial capital expenditure and operation and maintenance cost.

SCOPE OF RESEARCH

The objectives of this ongoing research effort are:

1. Establish the relative rates of nitrification for domestic wastewater treatment within an acclimated RBC fixed film system as a function of pH (between pH 6.0 and 9.0).

2. Observe and characterize the relative changes in the attached biological film as a function of pH.

3. Evaluate the efficacy of chemical addition to improve nitrification within an RBC fixed film system through the maintenance of an optimum pH.

4. Evaluate prime candidate chemicals for pH controlled nitrification for the RBC.

5. Develop economic design criteria for the RBC for pH controlled nitrification.

RESEARCH PLAN

RBC Pilot Facility

Secondary effluent from a trickling filter at the Penn State University (PSU) Wastewater Treatment Plant was selected for use in this evaluation. A schematic of the PSU wastewater treatment facility is shown in Figure 1. Approximately 1.7 m^3/day of wastewater which is recirculated around the PSU trickling filters is pumped to a pilot 0.5 m RBC pilot facility where temperature and flow are adjusted as desired. The character of the wastewater entering the RBC is described in Table 1. The pilot RBC facility is shown in Figure 2. Operating characteristics of the pilot RBC are presented in Table 2.

The operation of the RBC pilot unit is maintained at a hydraulic loading of 81 ℓ/m^2/day (2 gpd/ft^2), a rotational speed of 13 rpm, and a temperature in the first stage of 20°C. Preliminary pilot plant data which are presented in Table 1 demonstrate that the character of the wastewater entering the RBC is relatively low in BOD_5 and suspended solids. These data also show that some nitrification is occurring in the PSU sewage treatment plant. The pilot RBC system is functioning essentially as a separate stage nitrification process rather than a combined carbon oxidation-nitrification system.

TABLE 1. RBC INFLUENT WASTEWATER CHARACTERISTICS

BOD_5 (soluble)	-	6 mg/ℓ
Suspended Solids	-	18 mg/ℓ
Temperature	-	21°C
Flow	-	1.7 m^3/day
NH_3-N	-	14 mg/ℓ
NO_2^--N + NO_3^--N	-	3 mg/ℓ

Phase I - Nitrification as a Function of pH.

The initial phase of the research has been devoted to the establishment of an RCB system which is capable of nitrifying the natural wastewater containing approximately 290 mg $CaCO_3/\ell$ of alkalinity. Figure 3 shows a typical diurnal pattern of the ammonia-nitrogen entering the PSU wastewater treatment plant. The dampening of this diurnal pattern by primary treatment is illustrated in Figure 4 which also shows the diurnal pattern of ammonia-nitrogen within stage 1 of the pilot RBC. The ammonia-nitrogen removal performance of the pilot RBC from start-up is shown in Figure 5. In approximately 2 weeks after start-up, the RBC was producing an effluent with less than 0.05 mg NH_3-N/ℓ. Figure 6 demonstrates the degree of treatment provided during steady state operation. Obviously, at the current hydraulic loading rate of 81 ℓ/m^2/day, the first stage of the RBC is the most important and removes the greatest amount of ammonia nitrogen. Stages 2 and 3 provide additional ammonia removal; however, stage 4 provides no real additional treatment. The pH levels in stages 1 and 2 were 7.5 and 7.6 respectively. However, as the amount of ammonia oxidized in stages 3 and 4 declined, the pH in the later

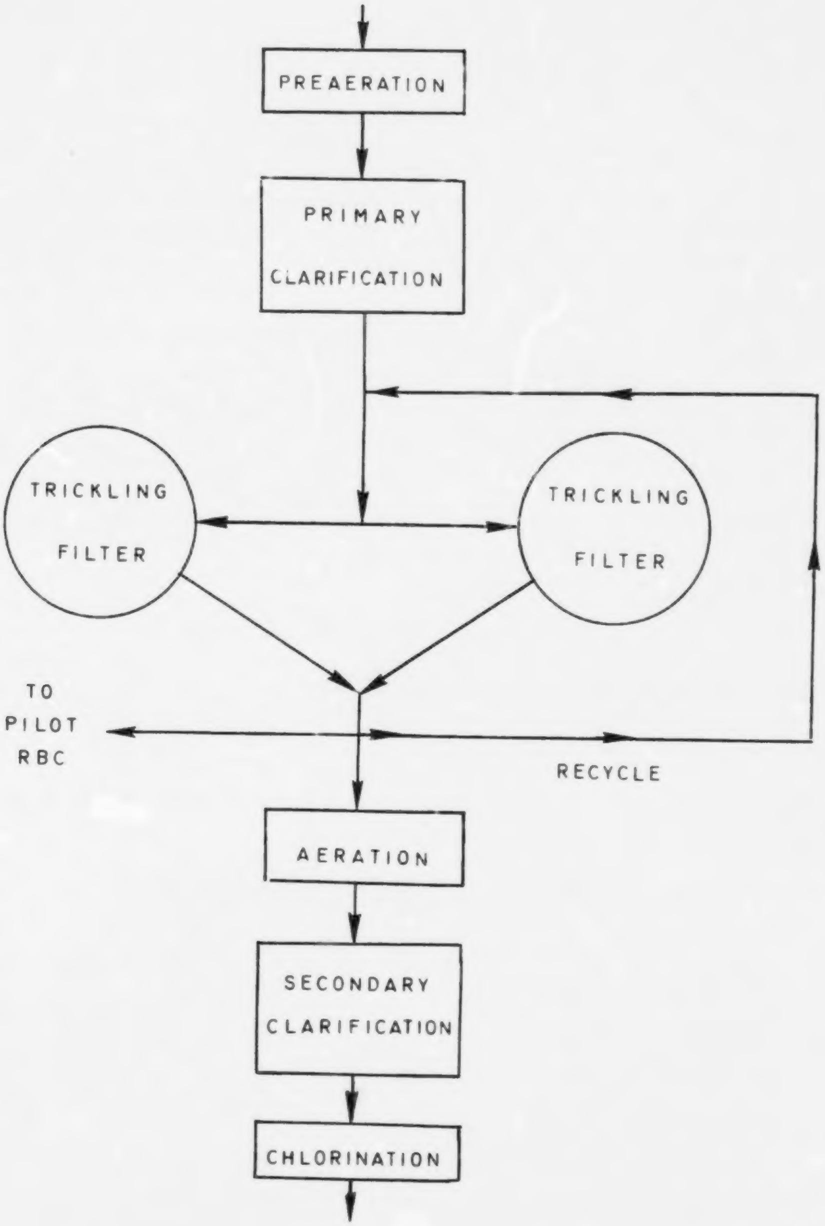
PREAERATION
PRIMARY CLARIFICATION
TRICKLING FILTER
TRICKLING FILTER
TO PILOT RBC
RECYCLE
AERATION
SECONDARY CLARIFICATION
CHLORINATION

Figure 1. Penn State University Wastewater Treatment Plant.

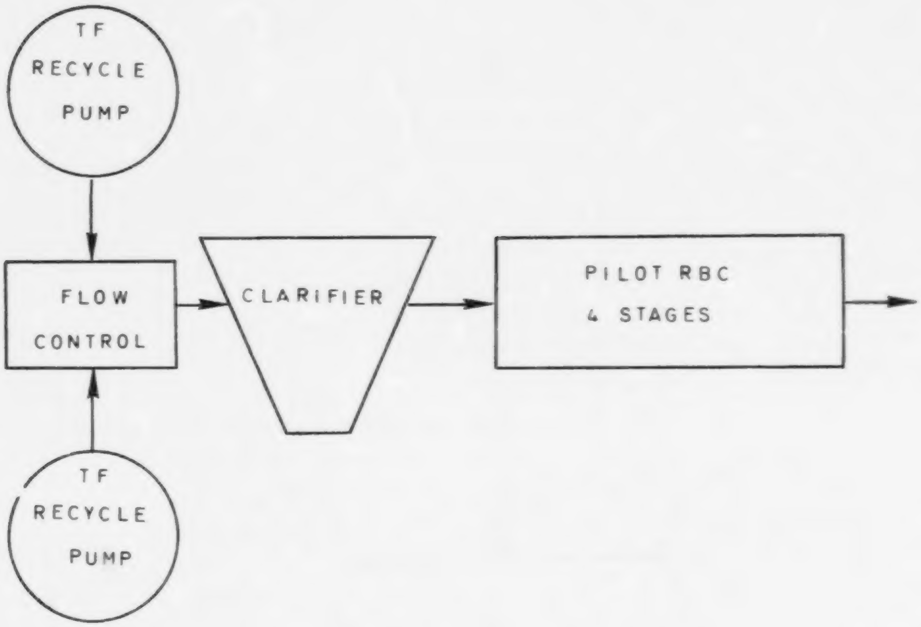
TF
RECYCLE
PUMP
FLOW
CONTROL
CLARIFIER
PILOT RBC
4 STAGES
TF
RECYCLE
PUMP

Figure 2. Pilot RBC Wastewater Treatment Plant.

stages rose due to the stripping of CO_2. The pH in stages 3 and 4 were 7.8 and 7.9, respectively.

TABLE 2. PILOT RBC OPERATING CHARACTERISTICS

Disc Diameter	-	0.5 m
Disc Area, Total	-	21.3 m^2
Number of Stages	-	4
Discs per Stage	-	9
Rotational Speed	-	13 rpm
Peripheral Speed	-	0.34 m/sec
Hydraulic Loading	-	81 ℓ/m^2/day
Hydraulic Residence Time	-	116 min.
Wastewater Temperature	-	20° ± 1°C

Based upon these preliminary data, subsequent phase 1 research will not utilize the entire RBC as a four stage treatment system. Instead, the influent flow will be increased four fold and split equally into each of the

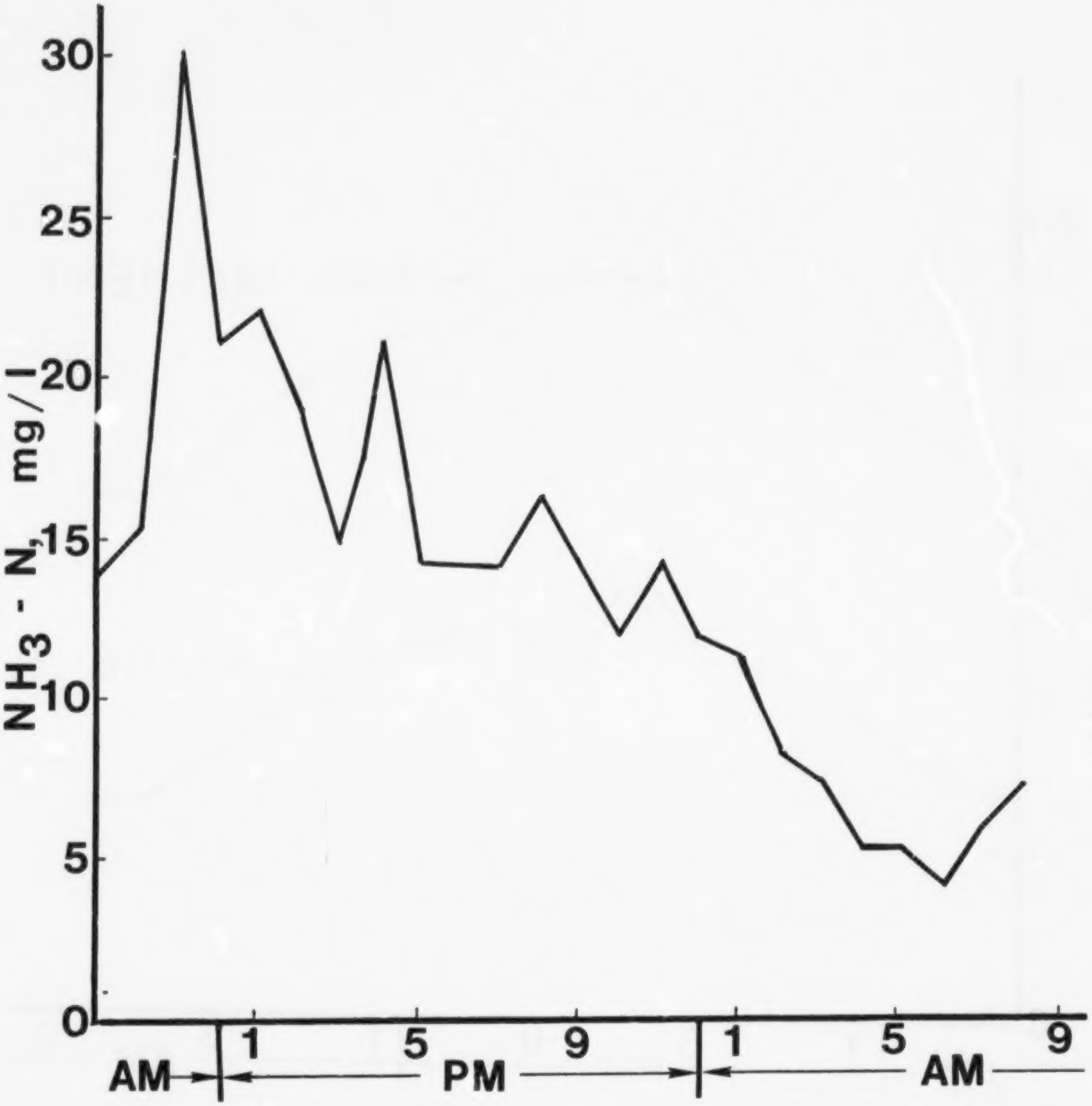

Figure 3. Diurnal Ammonia Variation Entering the Pennsylvania State University Wastewater Treatment Plant.

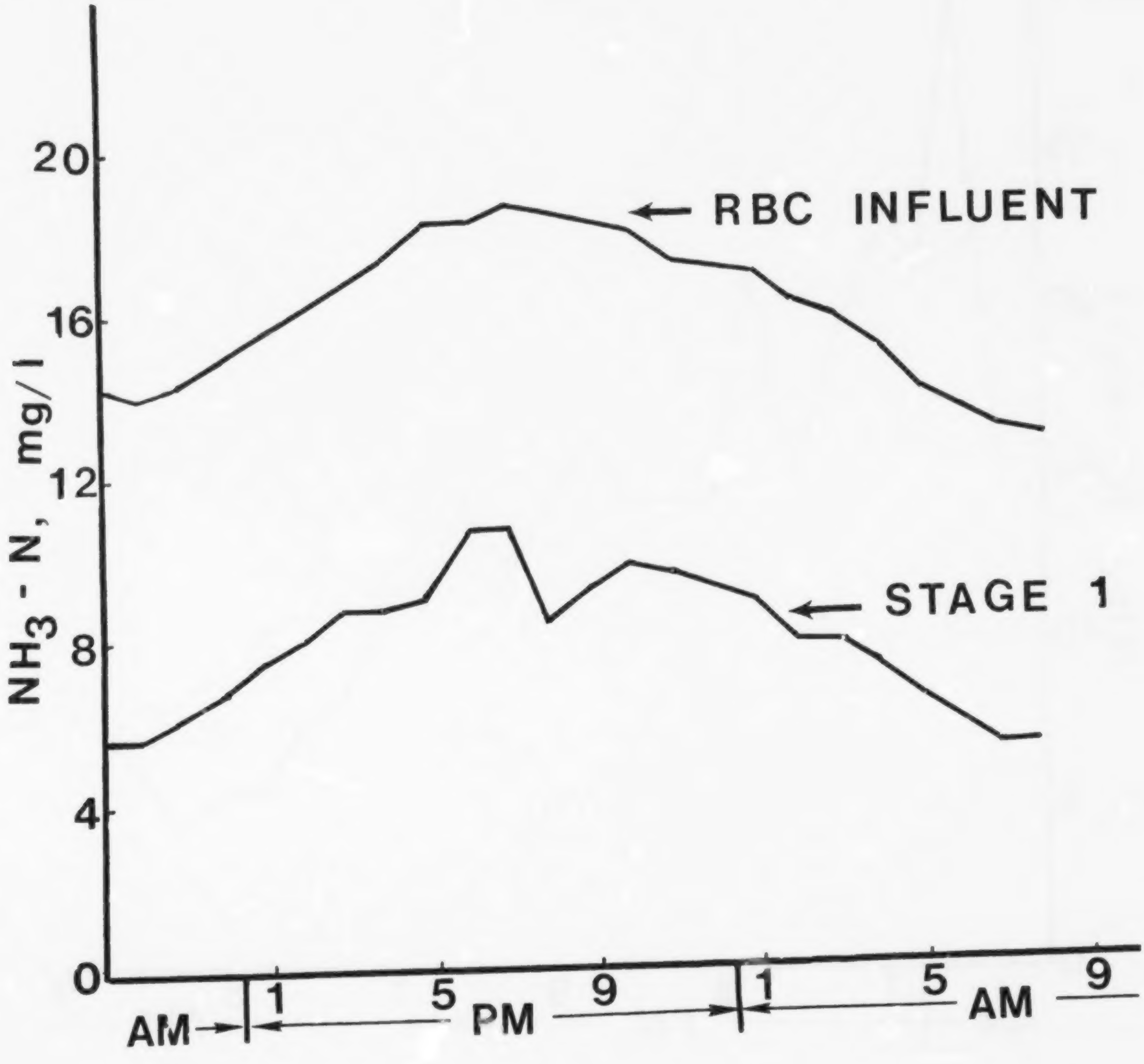

Figure 4. Diurnal Ammonia Variation Entering the RBC and in Stage 1 of the RBC.

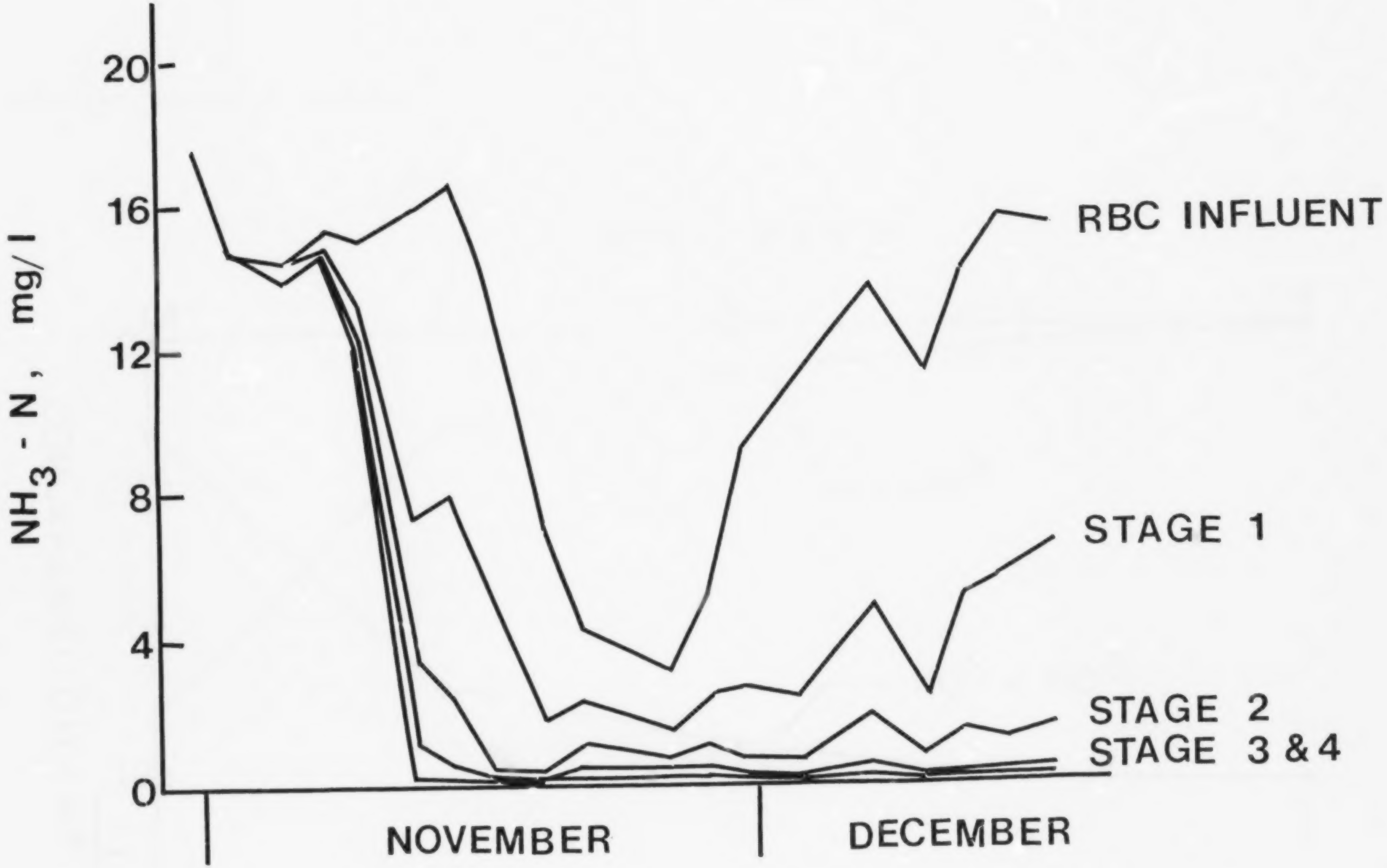

Figure 5. RBC Ammonia Removal Performance from Start Up.

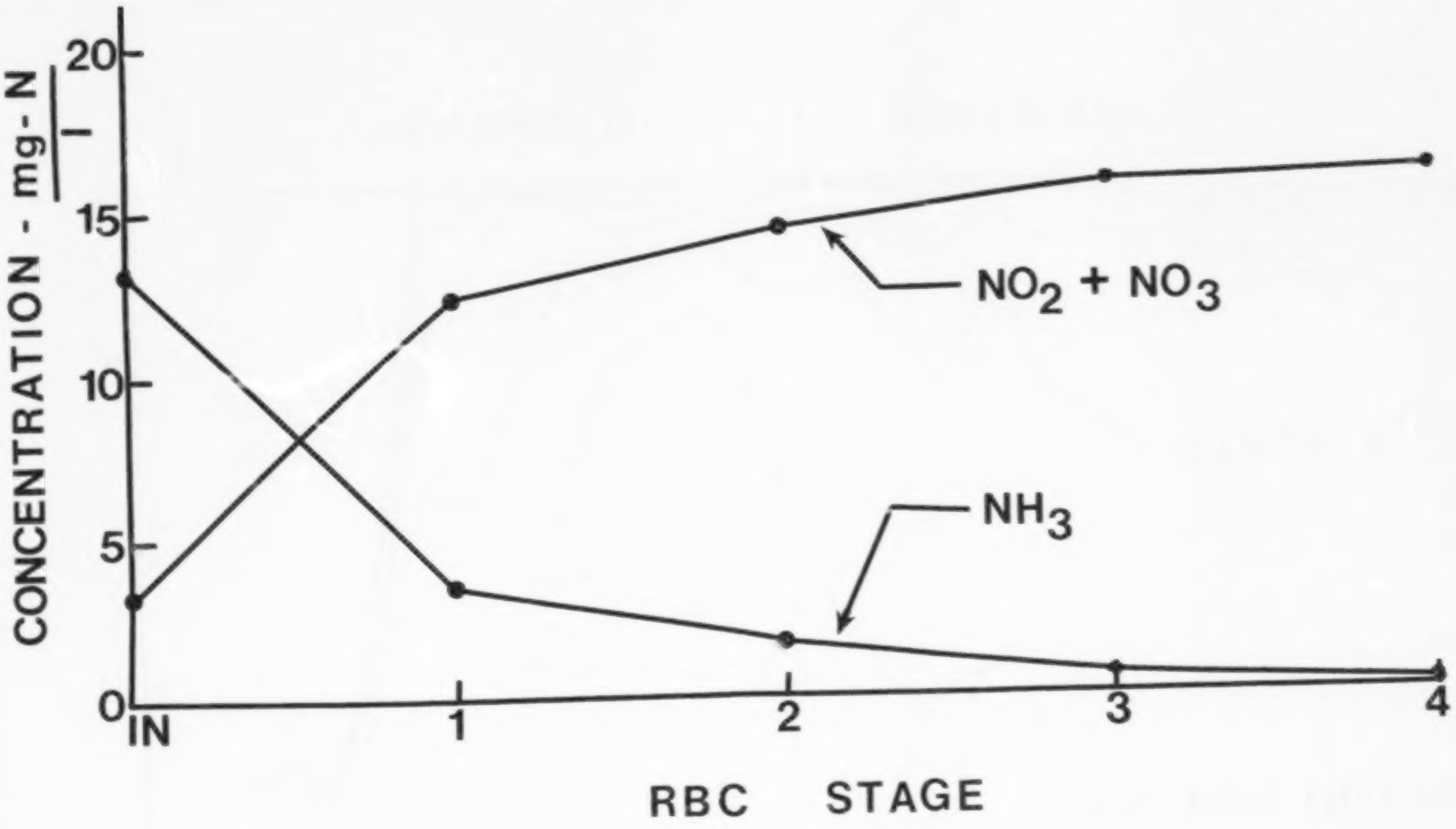

Figure 6. RBC Ammonia Conversion.

four stages. Therefore, each of the original stages will function as a single stage system. Because the first stage is the most critical and capable of adequately demonstrating the effect of pH, four different pH's and alkalinity conditions will be evaluated simultaneously. Each set of discs will receive the same wastewater, at the same temperature, the same influent dissolved oxygen level, and the same influent concentration of soluble BOD_5. Only the pH and alkalinity will be adjusted for each stage.

The pH adjustment phase will be used to examine RBC performance at pH 7.5, 7.0, 6.5 and 6.0. This part of the study will be terminated when steady state conditions are achieved. The experiment then will be repeated at pH 7.5, 8.0, 8.5 and 9.0. In each part, the pH 7.5 stage will serve as a control. The performance of the RBC stages will be evaluated on the basis of ammonia oxidation, solids generation, film establishment and population of nitrifiers on the fixed film surfaces. Each of the stages will start with a completely clean set of discs so that each stage will develop a fixed film nitrifying population which favors the particular pH of that stage. It is anticipated that differences in the nitrifying performance will be reflected in differences in the respective populations.

Biological Film Monitoring and Microbial Enumeration

The effect of altering the pH will result in changing the performance characteristics of the RBC system. This change will be reflected in the rate and magnitude of film buildup and the microbial composition of the disc film. The RBC pilot has been modified to facilitate the examination of the attached film. Each of the four stages is composed of eight molded polyethylene discs attached to a common central shaft. Each disc has approximately 0.63 m^2 of surface area available for film attachment. An additional flat plexiglass disc has been added to each state thereby providing an additional area of 0.27 m^2. Sections of Mylar are attached to each of the plexiglass discs to allow for easy removal. The biological films are removed from the Mylar and broken up by blending and evaluated for ammonia oxidizing and nitrite oxidizing bacteria as well as for total and volatile solids. The Nitrosomonas are grown in an ammonium-calcium carbonate media and the Nitrobacter are grown in a nitrite-calcium carbonate media (16). The method of enumeration utilizes a 96 well tissue culture plate (17,18). Because of the slow growth rate of the organisms the microbial enumeration procedure requires approximately 28 days of incubation before the final MPN values can be determined. Preliminary data at pH 7.5 which demonstrates the biological slime buildup and microbial enumerations are provided in Figures 7 and 8, respectively. The biological film development continued throughout the seven weeks of the preliminary operation of the pilot RBC. The heaviest growth was noted on the first stage with decreasing growth in the later stages. Although the nitrification performance of the unit had stabilized by mid-November, the film continued to develop throughout the balance of the initial test period. This observation indicated that only a fraction of the biological film was actively engaged in ammonia and nitrite oxidation. The enumeration of ammonia oxidizing and nitrite oxidizing bacteria as shown in Figure 8 was performed after seven weeks of operation. In stages 1 and 2, the number of Nitrosomonas bacteria were greater than Nitrobacter bacteria; however, this trend was reversed in stages 3 and 4. This change apparently was due to the disappearance of ammonia-nitrogen in stages 3 and 4 and lingering concentration of nitrite-

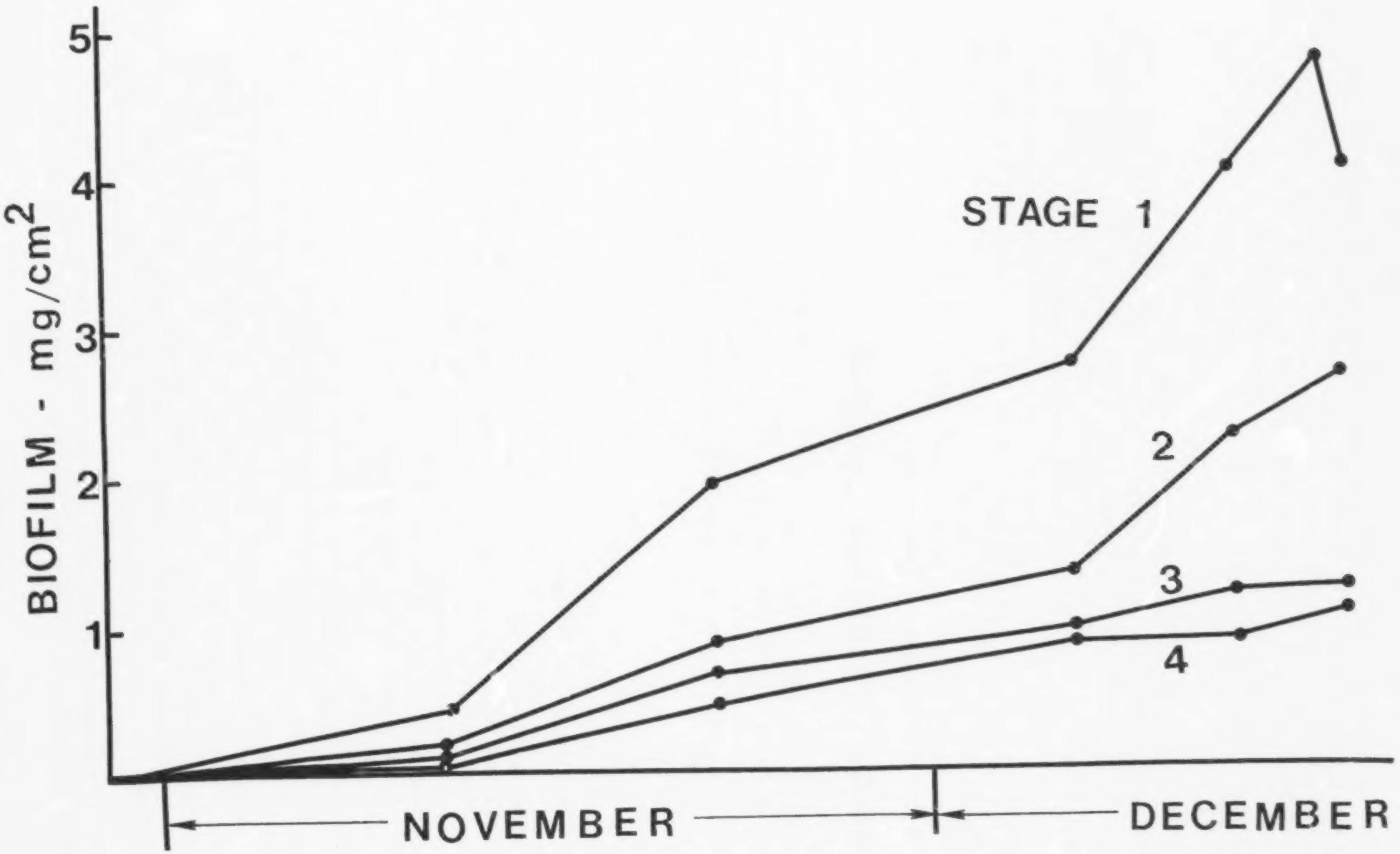

Figure 7. RBC Biofilm Development with Time.

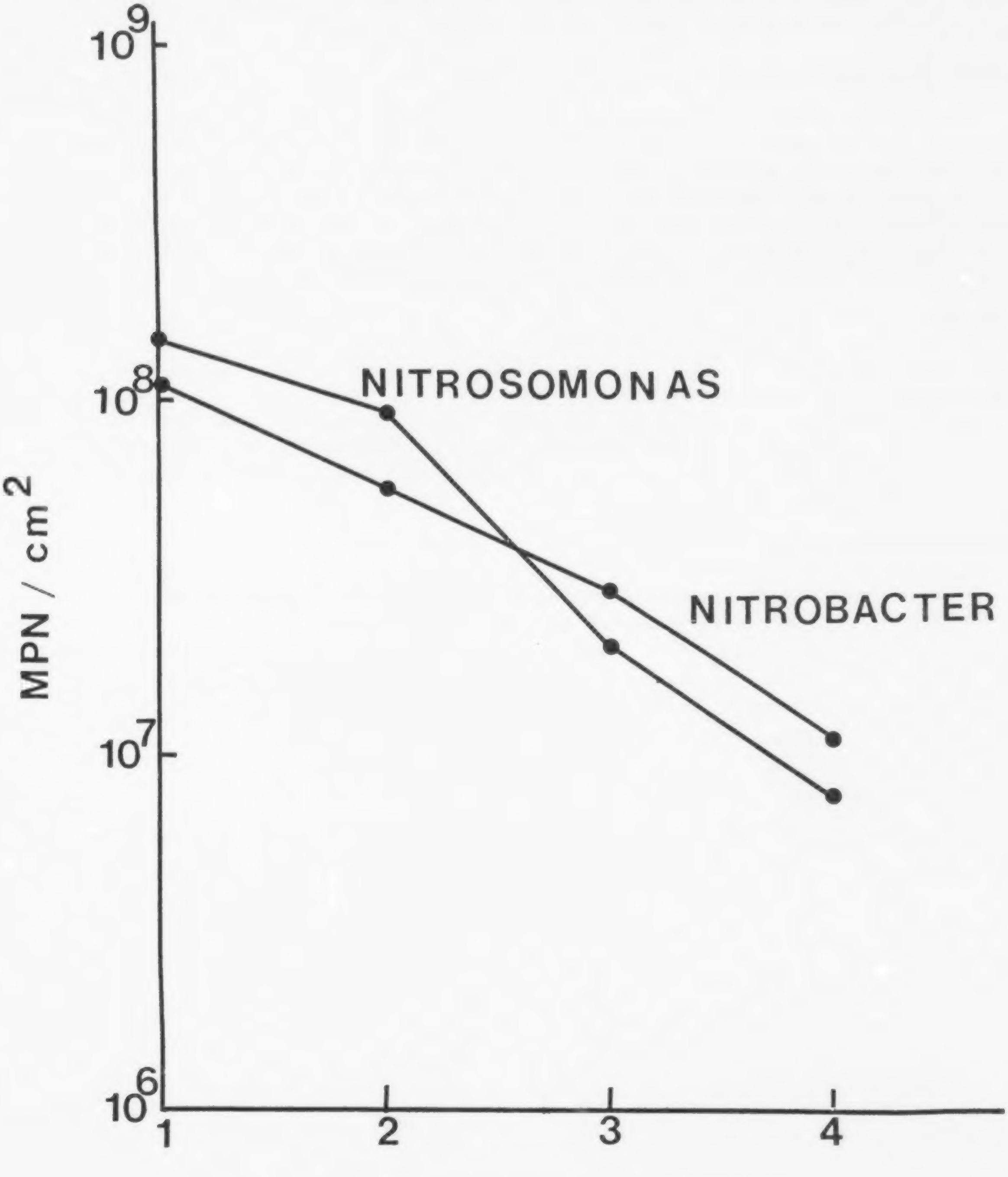

Figure 8. RBC Microbial Enumerations.

nitrogen. Both populations decreased by approximately an order of magnitude across the four stages of the RBC pilot plant.

Phase II - pH Adjustment

The last phase of this research effort will be to examine alternative schemes for pH control so that nitrification can be optimized. After phase I is completed, the optimum pH level will be identified. The wastewater will be altered through chemical addition to simulate a low pH-low alkalinity wastewater. Alternative chemical additives will be evaluated in enhancing the nitrification process. The alternative chemicals to be evaluated are lime, soda ash, sodium bicarbonate, and sodium hydroxide.

SUMMARY

The results of the two year research effort will be used to provide more definitive guidance on the long and short term effect of pH for fixed film RBC nitrification. The results also will more clearly establish the optimum pH level required for nitrification with full scale RBC facilities as well as the means by which proper pH adjustment can be made to insure optimum nitrification.

ACKNOWLEDGMENT

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NITRIFICATION OF MINICIPAL WASTEWATER USING ROTATING BIOLOGICAL CONTACTORS

by

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INTRODUCTION

The need for nitrogen control has been realized in recent years to upgrade the quality of the waters receiving wastewater effluents. One method of upgrading wastewater effluents is using biological nitrification within the treatment process.

The biological process in which ammonia nitrogen is converted to nitrate is a two-step process in which nitrosomonas bacteria convert ammonia to nitrate and nitrobacter and then convert the nitrite nitrogen to nitrate nitrogen. The stoichiometric equations for this synthesis of oxidation process can be summarized as follows (1):

$$55\ NH_4^+ + 76\ O_2 + 109\ HCO_3^- \xrightarrow{\underline{\text{Nitrosomonas}}} \underset{\text{cells}}{C_5H_7NO_2} + 57\ H_2O + 104\ H_2CO_3$$

$$400\ NO_2^- + NH_4^+ + 4\ H_2CO_2 + HCO_3^- + 195\ O_2 \xrightarrow{\underline{\text{Nitrosobacter}}}$$
$$C_5H_7NO_2 + 3\ H_2O + 400\ NO_3^-$$

Since carbonic acid is formed in this process, and bicarbonate levels are reduced, as bicarbonate alkalinity is consumed the pH will drop. It can be theoretically calculated that approximately 7 mg/l of alkalinity will be destroyed for every milligram of ammonia nitrogen that is oxidized.

Presently, Rotating Biological Contactors (RBC's) are being used for the secondary treatment of domestic wastewater, industrial wastewater, for septage treatment, and for mitrification and denitrification. There are many firms manufacturing and marketing such devices across the nation. Differences exist between the manufactured units currently available. Such differences as surface configuration, materials, sizes and the surface to volume ratio of the reaction chambers exist in various designs and applications.

The design considerations used in the selection of RBC's are as follows:

1. Influent wastewater characteristics
2. Hydraulic loading
3. Operation temperature
4. pH control
5. Effluent wastewater requirements
6. Disc velocity
7. Size and surface area of media
8. Substrate Loading rates

It is important that these parameters be investigated carefully before selection or sizing the units.

Most of the available literature to date deals with reduction of BOD_5 within an RBC unit. Limited data has been published on the performance of RBC units when used to achieve biological nitrification of a domestic wastewater.

The primary objectives of the research study presented in this paper were:

1. Perform a comparative analysis of the nitrification efficiencies of the rotating biological contactor system with an operating two-stage activated sludge system.
2. Establish treatability parameters involved with an RBC system. These parameters include organic and nitrogen loading rates, sludge production, surface area requirements for proper F/M ratios.

EXPERIMENTAL FACILITIES AND PROCEDURES

Both laboratory bench scale units and field pilot units were used for data collection.

The laboratory study was conducted in the Environmental engineering laboratories located at Northeastern University. The bench scale units used in this study consisted of two pairs of contactors with a surface area equal to 9.08 ft^2/stage. There were ten (10) discs/stage. Volumes were equal to .7 gal/stage. These units were constructed at Northeastern University using plastic media supplied by EPCO-Hormel.

The wastewater flow scheme is shown in Figure 1. The units were fed secondary treated effluent from the Marlboro Easterly Wastewater Treatment Plant. Two units (B and D) received normal secondary effluent while two units (A and C) received the same effluent but the feed was "spiked" with ammonia chloride so that the ammonia concentration was 40 mg/l NH_3-N. Fresh effluent was transported to the laboratory every other day.

'Units A and B were four stage units with a total surface area of 36.3 ft^2 and liquid volume of 2.8 gal. Units C and D were single stage units with an expanded liquid volume chanber. Throughout the study, the effective liquid volume of Units C and D was 2.8 gal.

Grab samples were taken at intervals, based on flow and detention time per stage, so that the flow could be followed through the units.

Figure 2 shows the set-up for both a single stage unit and a four stage unit.

Four different pilot RBC systems were used during the field portion of this study. The units were located at the Marlboro Easterly Wastewater Treatment Plant in Marlboro, Massachusetts. Each of the units received effluent pumped directly from the secondary clarifier of the plant's activated sludge treatment system. The pilot units were located in the treatment plant's underground service tunnels.

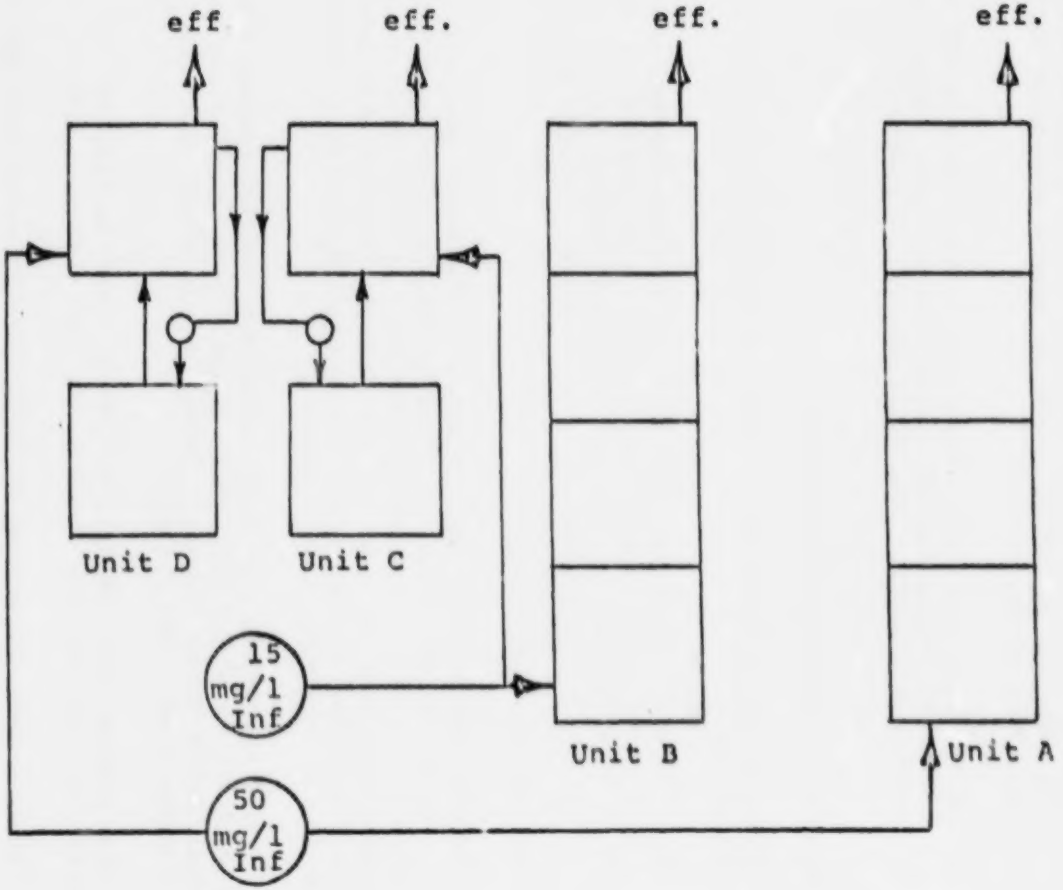
eff
eff.
eff.
eff.
Unit D
Unit C
Unit B
Unit A
15 mg/1 Inf
50 mg/1 Inf

FIGURE 1: FLOW SCHEME FOR LABORATORY NITRIFICATION STUDY

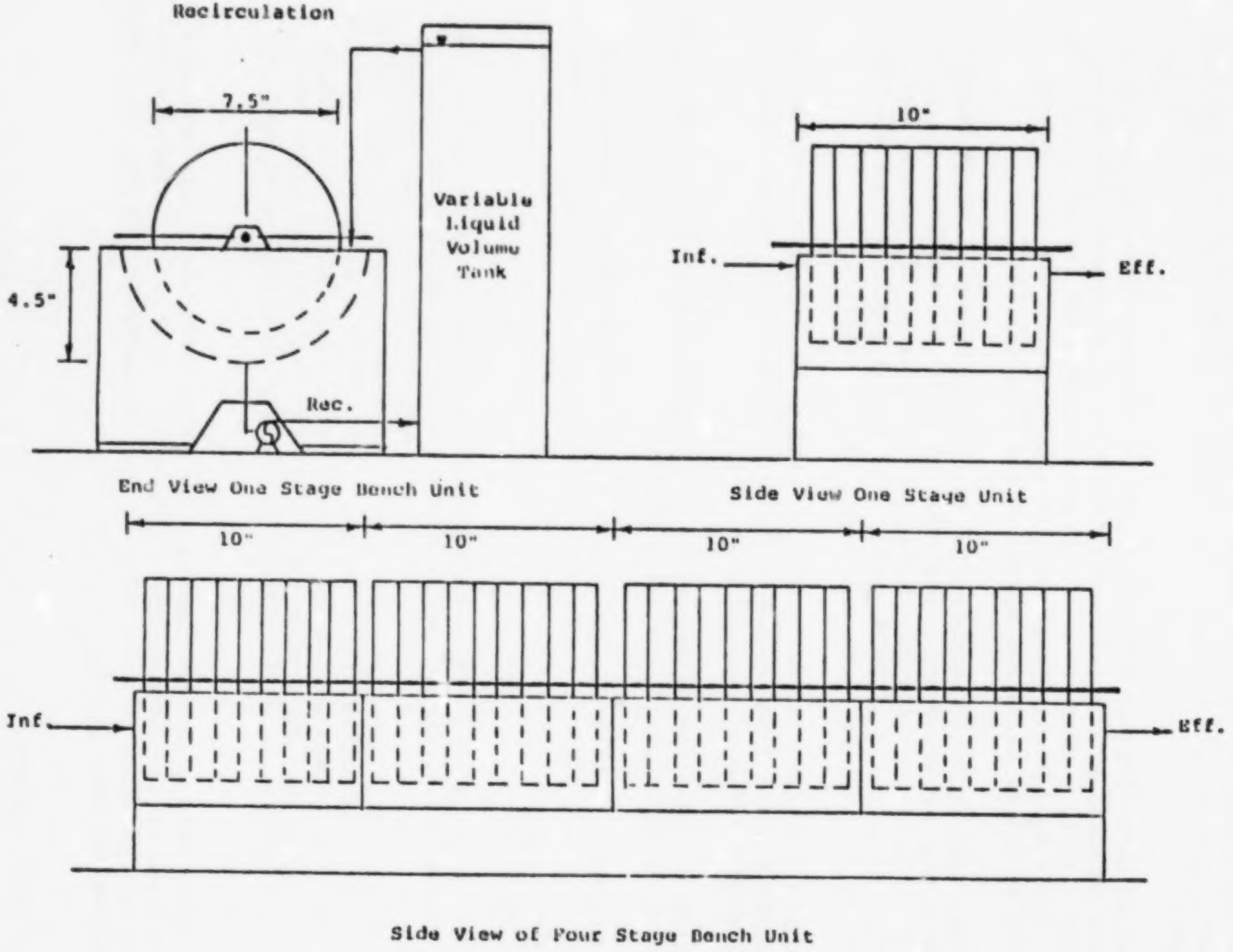
Recirculation
7.5"
4.5"
Variable Liquid Volume Tank
Rec.
End View One Stage Bench Unit
10"
Inf.
Eff.
Side View One Stage Unit
10"
10"
10"
10"
Inf.
Eff.
Side View of Four Stage Bench Unit

FIGURE 2: Laboratory Set-Up for Both Single and Four Stage RBC Units Used for Nitrification Evaluation

The first system (Unit 1) utilized was a four ft. diameter four stage system supplied by the Environmental Pollution Control (EPCO), a subsidiary of the George A. Hormell Company of Austin, Minnesota. There was 1,570 square feet of disc surface area, equally spaced in each stage. The unit was run with a liquid volume of 150 gallons. The volume to surface ratio of the unit was 0.095 gal/ft^2.

The second unit (Unit 2) utilized a two ft. diameter four-stage contactor which was supplied by the Autotrol Corporation of Milwaukee, Wisconsin. This unit had a total surface area of 250 square feet, the liguid volume was 27 gallons, there were 36 bio-discs equally spaced within the four stages, and a surface to volume ratio was approximately 0.12 $gal/ft.^2$

The third unit (Unit 3), constructed by the Department of Civil Engineering at Northeastern University, consisted of a two ft. module containing 40 discs, 10 discs per stage with a total surface area of 327 square feet. The discs were constructed of EPCO media. The unit's liquid volume was 129 gal., resulting in a volume to surface ratio of 0.4 gal/ft^2.

The fourth unit (Unit 4) was a single stage unit, constructed by Northeastern University with a surface area of 169 ft^2 and liquid volume equal to 12.4 gal. giving a ratio of 0.07 gal/ft^2.

Operation of the units varied, with Unit 2 being run as both a four stage unit, and also as two separate two stage units.

Grab samples were taken at intervals, based on flow and detention time. During the study, all ammonia values are average values based on three samples taken at three consecutive detention times. Analyses were completed within 24 hours of sampling.

The nitrification pilot units were operated a total of seven months, beginning in September and ending in December of 1976; and again from April through June of 1977. The first time period saw Unit 1 run as two separate two stage units. One of the two stage units acted as a "control" unit and did not receive additional alkalinity addition. All other units received additional alkalinity. Unit 2 was run as a four stage unit, and Unit 4 as a single stage unit. During the second time period of the nitrification pilot study, Units 1, 2, and 3 were run as four stage units, all receiving additional alkalinity.

All samples were grab type samples. Analyses were either run immediately after collection, or the samples were refrigerated to 4°C. All analyses were completed within 24 hours of collection on the refrigerated samples. All water quality testing followed Standard Methods (2).

BENCH SCALE STUDY RESULTS

The nitrification bench study was run for approximately two months following a one month start-up period. The influent to Unit B (4-stage) and Unit C (single stage) averaged between 10 to 15 mg NH_3-N/1 for the remainder of the study. The ammonia influent and effluent concentrations for all of the units are shown in Figures 3 and 4.

Both Unit B and Unit C had effluent ammonia nitrogen concentrations less than 1.0 mg/l (with the execption of one data point) throughout the study period. Unit C had a slight increase in ammonia nitrogen in the effluent following increases in influent ammonia concentrations. However, during periods of constant loading rates, the ammonia nitrogen concentration of the units was less than 0.1 mg NH_3-N/l.

Unit A was able to maintain low ammonia effluent concentrations (less than 0.3 mg NH_3-N/l) during May. The loading rates during this period were low and relatively stable. Unit C (single stage) followed a similar trend during the same time period.

The increased unit loading in early June resulted in high ammonia concentrations in the effluents, as a result of this shock load. Both units recovered from the shock loading increase, however, the four stage unit did recover more rapidly. The four stage unit (Unit A) also achieved a higher degree of treatment of the wastewater during this portion of the study. This early June loading increase in Units A and D resulted from increased ammonia concentration only while the hydraulic loading was fairly constant.

The peak in ammonia effluent concentrations which occurred in the middle of May was associated with increased hydraulic loading rates on the RBC units. The hydraulic residence time per stage of unit was reduced from approximately 3.0 hours/stage to 1.5 hours/stage. This hydraulic loading increase also caused the daily ammonia loading rate to double. The two four-stage units had little change in effluent quality while both single stage units (Units C and D) had higher ammonia effluent concentrations. Since all of the units had the same liquid volume, the results indicate that staging will help reduce the effects of a simultaneous increase in flow and ammonia concentration.

Ammonia Loading Rates and Nitrate Appearance

The ammonia nitrogen loading rates to each unit were calculated and the results show in Figure 5. The loading rates are expressed as lbs. NH_3-N applied/day-1000 ft^2. The lbs. NH_3-N converted/day-1000 ft^2 was also calculated. The data shown in Figure 5 excludes data from periods just after sharp loading changes, and as a result, represents data recorded during periods of stable operation. The mathematical relationship for lbs. of ammonia nitrogen applied vs. lb. ammonia nitrogen converted is shown on Figure 5. This equation was calculated by the "least squares method", for loadings up to 0.8 lbs. NH_3-N applied/1000 ft^2-day. The correlation coefficient of the data was 0.99. Most of the ammonia nitrogen applied to the RBC units at such loading rates was converted nitrate nitrogen.

The conversion of ammonia nitrogen represents nitrogen which was converted to nitrate nitrogen by the nitrifying bacteria plus ammonia nitrogen converted into cellular nitrogen compounds associated with new biomass production.

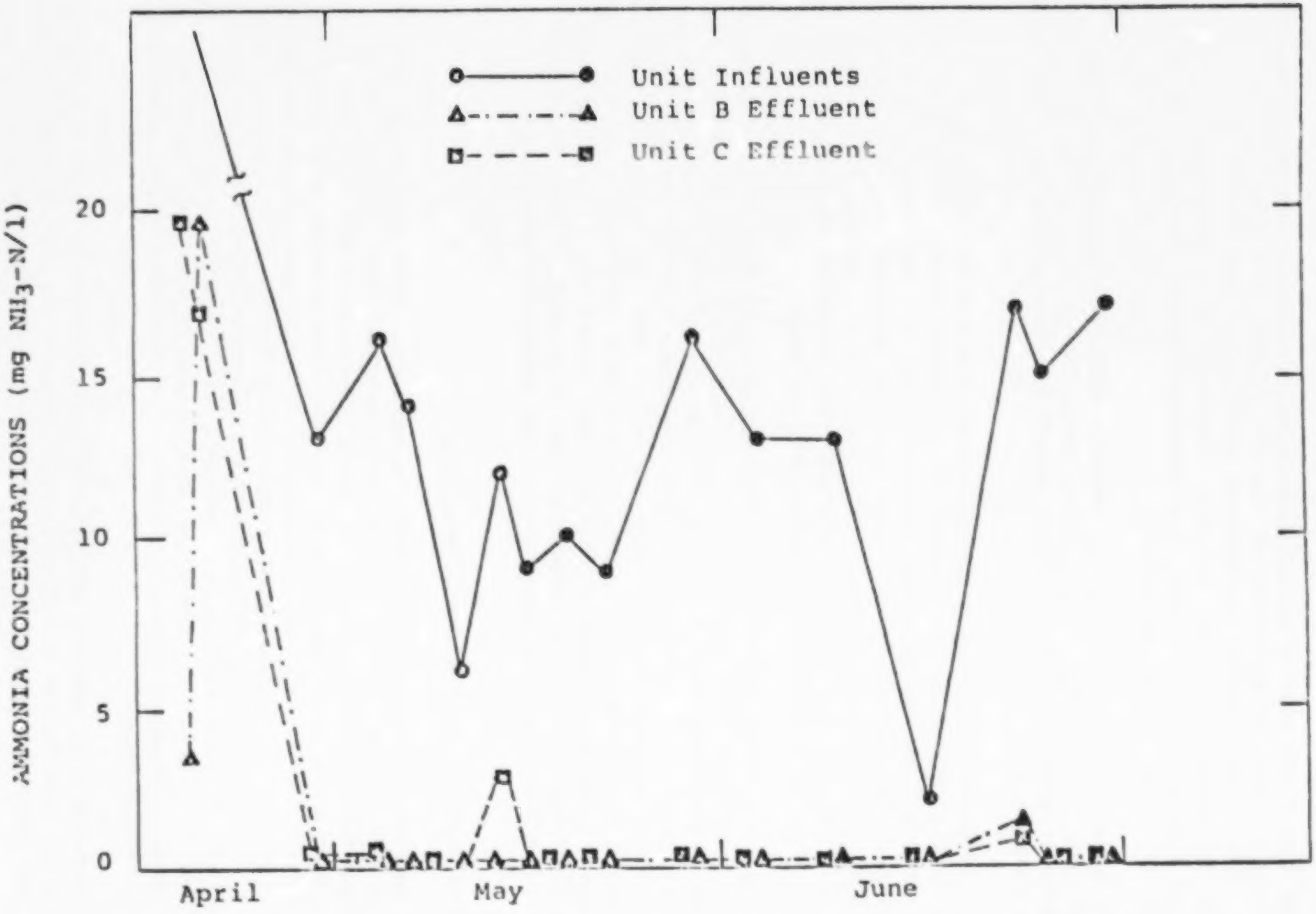

FIGURE 3: AMMONIA NITROGEN INFLUENT AND EFFLUENT CONCENTRATIONS
(Bench Scale Units B and C)

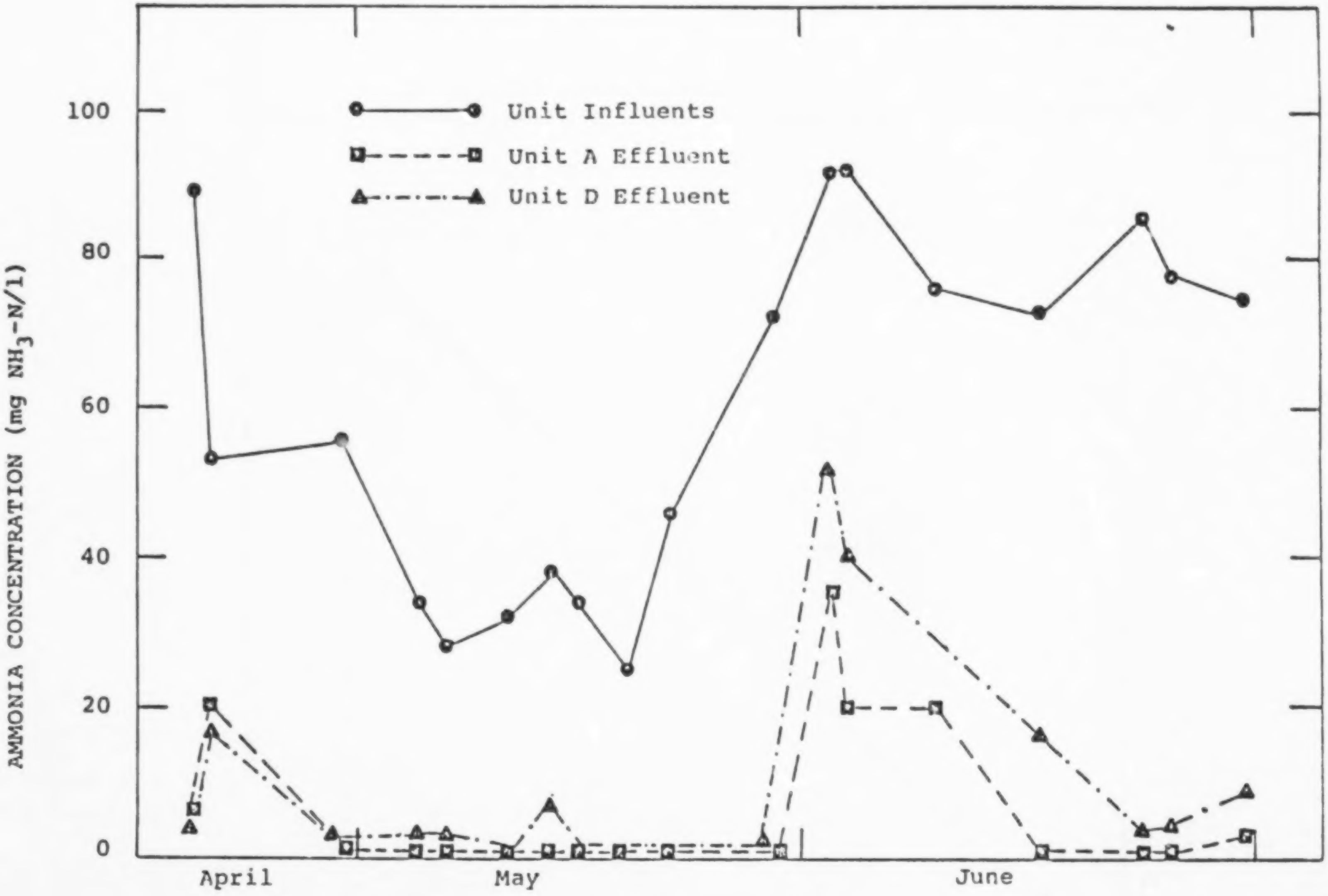

FIGURE 4: AMMONIA NITROGEN INFLUENT AND EFFLUENT CONCENTRATIONS (Bench Scale Units A and D)

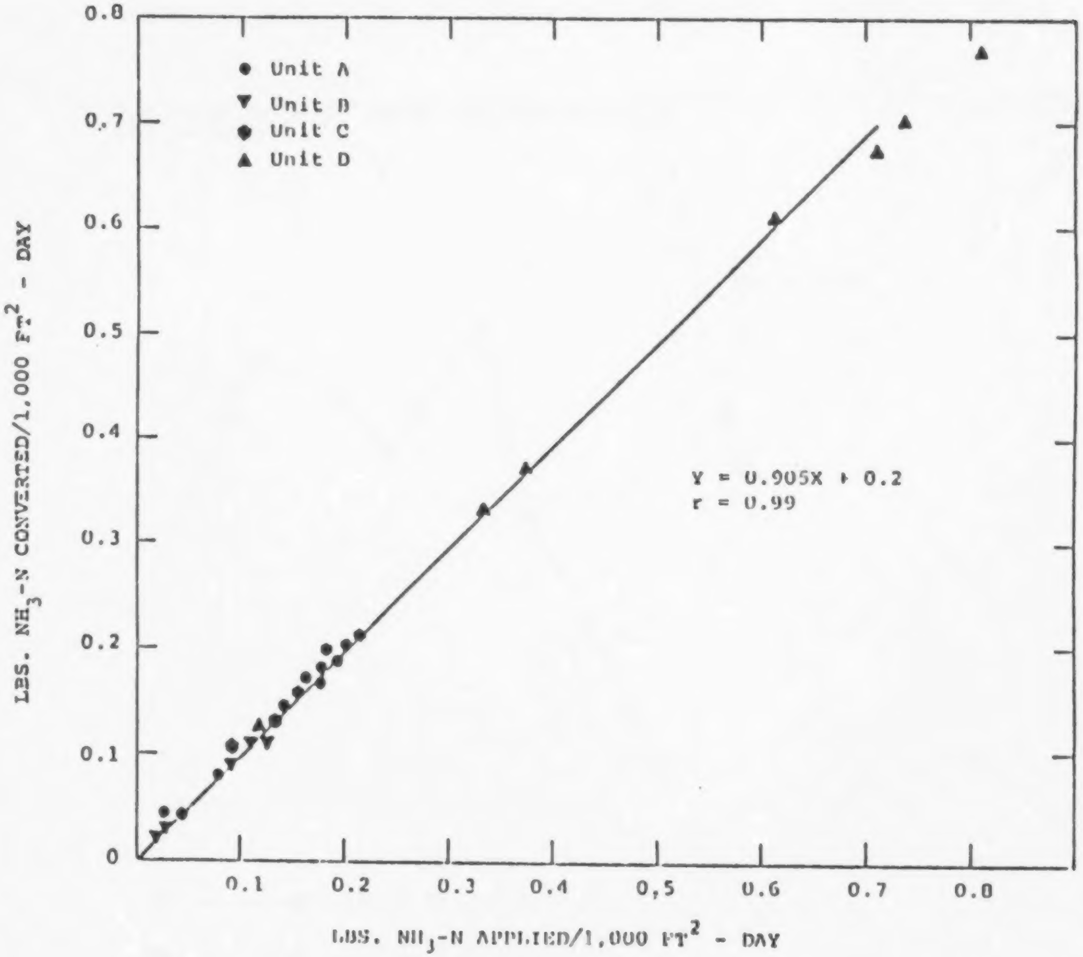

FIGURE 5: Pounds Ammonia Nitrogen Converted to Nitrate Nitrogen vs. Pounds Ammonia Applied

Table 1 represents a summary of the ammonia nitrogen conversion and the nitrate nitrogen appearance data for this study period. An average of 93 per cent of the ammonia nitrogen which was removed from the influent feed appeared as nitrate nitrogen in the treated effluents. This data indicates that most of the converted ammonia nitrogen is nitrate nitrogen, a result of nitrification, and will remain in the effluent. Only seven percent of ammonia mass balance appeared to be associated with new biomass.

In caluclating the mass balance, when nitrate appearance was higher than ammonia conversion, ammonia was assumed to be totally converted to nitrate.

TABLE I: PERCENT OF CONVERTED AMMONIA NITROGEN WHICH APPEARED AS NITRATE NITROGEN.

UNIT A	95%
UNIT B	98%
UNIT C	88%
UNIT D	90%
MEAN VALUE FOR ALL UNITS	93%

The percent ammonia nitrogen converted is plotted against ammonia nitrogen loading rate in Figure 6. In all cases, the four stage units converted 95 percent of the ammonia nitrogen. The maximum loading rate for four stage units was 0.2 lbs. NH_3-N/1000ft^2-day. The single stage units had 95 percent ammonia nitrogen conversion for loadings up to 0.8 lbs. NH_3-N/1000ft^2-day, with two exceptions.

These efficiencies represent steady state periods of operation, without shock loading effects. Higher loading rates could not be reasonably evaluated in the laboratory study due to the logistical problems involved with transportation of the influent.

Alkalinity and pH

Additional alkalinity was added to the influent to insure that sufficient inorganic carbon would be available for nitrification. Alkalinity additions resulted in influent alkalinity concentrations of 250 to 350 mg/l as $CaCO_3$ for Units B and C. Units A and D which has higher ammonia nitrogen concentration in the influent had average alkalinity concentrations of approximately 500 mg/l as $CaCO_3$. Effluent alkalinity was greater than 150 mg/l as $CaCO_3$ in all cases.

The pH values ranged between 8.0 and 9.0. Both Na_2CO_3 and $NaHCO_3$ were used as a buffer and also to supply inorganic carbon. Influent pH values remained constant over two to three week periods, and there were not any sudden pH changes over the study period. The average pH value was 8.5 for influents during May.

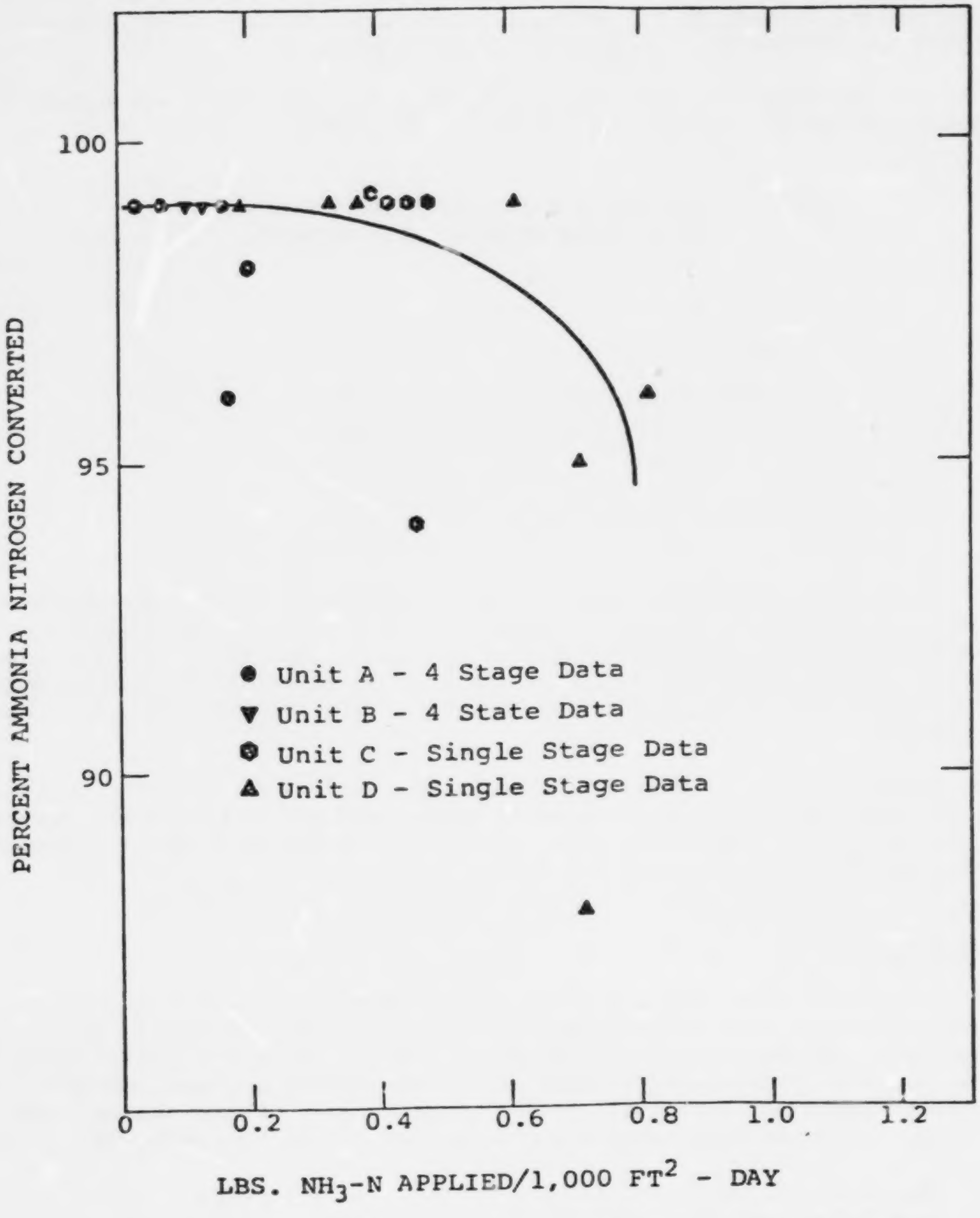

FIGURE 6: PERCENT AMMONIA NITROGEN CONVERTED TO NITRATE NITROGEN VS. AMMONIA APPLIED (BENCH SCALE UNITS)

Organic Nitrogen

Throughout the study, influent organic nitrogen concentrations averaged 4.4 mg N/l, while effluent concentrations averaged 1.2 mg N/l.

Units A and B, the four stage units, had less organic nitrogen in their effluent samples than did the single stage units. All units averaged less than 2.0 mg N/l in the effluent samples, while the four stage units averaged organic nitrogen concentration below 1.0 mg N/l. Complete data for this portion of the study is given in O'Shaughnessy and Blanc (3).

Staging

Several times during the study, samples were taken in each of the four stage units.

The results indicated that most of the nitrification occurred in the first stage. In all cases, at least 90 percent of the ammonia nitrogen conversion occurred in the first stage. The average first stage ammonia nitrogen conversion was 95 percent of the ammonia applied. The data indicated that for steady state condition, most of the nitrification occurred in the first stage.

PILOT STUDY RESULTS

The pilot studies conducted at Marlboro Easterly W.T.P. were run over two separate time periods. However, discussion of all of the results will be combined in this section. The two separate time periods were Fall, 1976, and Spring, 1977. Wastewater temperatures in the units were in the same range (10°C to 16°C) for both study periods.

Fall Study Results

During the 1976 study period, four separate units were run. They were respectively:

1. N.U. single stage unit - 0.07 gal/ft^2;
2. EPCO-HORMEL two-stage unit - 0.09 gal/ft^2 (used as a control unit without any alkalinity additions);
3. EPCO-HORMEL two stage unit - 0.09 gal/ft^2;
4. AUTOTROL four stage unit - 0.12 gal/ft^2.

Data from the study indicated that nitrification of secondary treated municipal wastewater was incomplete unless additional alkalinity was added. Alkalinity additions were always directly to the first stage of a unit, except in the case of the control unit.

Figure 7 is a plot of data from the two separate two stage units. The unit receiving alkalinity additions averaged 91 percent ammonia nitrogen conversion over a two-month period, while the identical control unit only averaged 22 percent ammonia conversion. The influent alkalinity over the two month period averaged 80 mg as $CaCO_3$/1, and the pH averaged 7.0. An average of approximately 185 mg $CaCO_3$/1 of alkalinity was required to carry out the biological oxidation of the ammonia nitrogen. The poor ammonia conversion ratio of the control unit when compared to the other two stage unit indicated that alkalinity addition was required for a high level of nitrification to be completed.

Ammonia Loading Rates

Figure 8 shows the loading rate and effluent ammonia concentrations for the four stage RBC unit. As in the bench scale study, the ammonia effluent concentrations followed the loading rate. Sudden increases in ammonia loading rate resulted in increased ammonia nitrogen concentrations in the unit's effluent. For all units, the lowest effluent ammonia concentrations occurred when a constant loading rate below 0.30 lbs NH_3-N/1000ft^2 - day was maintained. The sharp increase in effluent ammonia during early December can be attributed to a sudden drop in temperature below 10°C.

The conversion of ammonia in both the two stage unit and the first two stages of the four stage unit was fairly constant (Figure 9). An average of 90 percent of the applied ammonia was converted for a loading rate of 0.40 lbs NH_3-N/1000ft^2-day. Loading rates of 0.20 lbs NH_3-N/1000ft^2-day and less produced between 94-99 percent ammonia nitrogen conversion (Figure 9). There was a noticeable reduction of the percent ammonia converted as the ammonia loading rate was increased. Only 80 percent ammonia conversion was recorded at a loading of 0.5 lbs NH_3-N/1000ft^2-day. The data plotted in Figure 9 represent loading rates and conversion of ammonia within two stage units.

The first stage loading rates and conversion efficiency was calculated for each unit. Neglecting shock load and start-up data, the following average ammonia conversion percentages were calculated for the study period:

Single stage unit - 56 percent at 0.51 lbs NH_3-N/1000ft^2-day,

2-stage unit - 65 percent at 0.61 lbs NH_3-N/1000ft^2-day,

4-stage unit - 73 percent at 0.78 lbs NH_3-N/1000ft^2-day.

While there was a large range in both ammonia nitrogen loading rates, the majority of ammonia nitrogen was converted in the first stage.

The data plotted in Figure 10 represents ammonia nitrogen conversion and loading rates for both two and four stage units. An average of 90 percent of the applied ammonia nitrogen was converted for loading rates up to 0.6 lbs NH_3-N/1000ft^2-day. This data includes both two and four stage units. For periods of relatively steady state loading, the four stage unit averaged 96 percent ammonia conversion, while the two stage unit averaged 91 percent ammonia conversion. Based on this data, the four stage unit appeared to achieve better ammonia nitrogen conversion than the two stage unit. Since each utili-

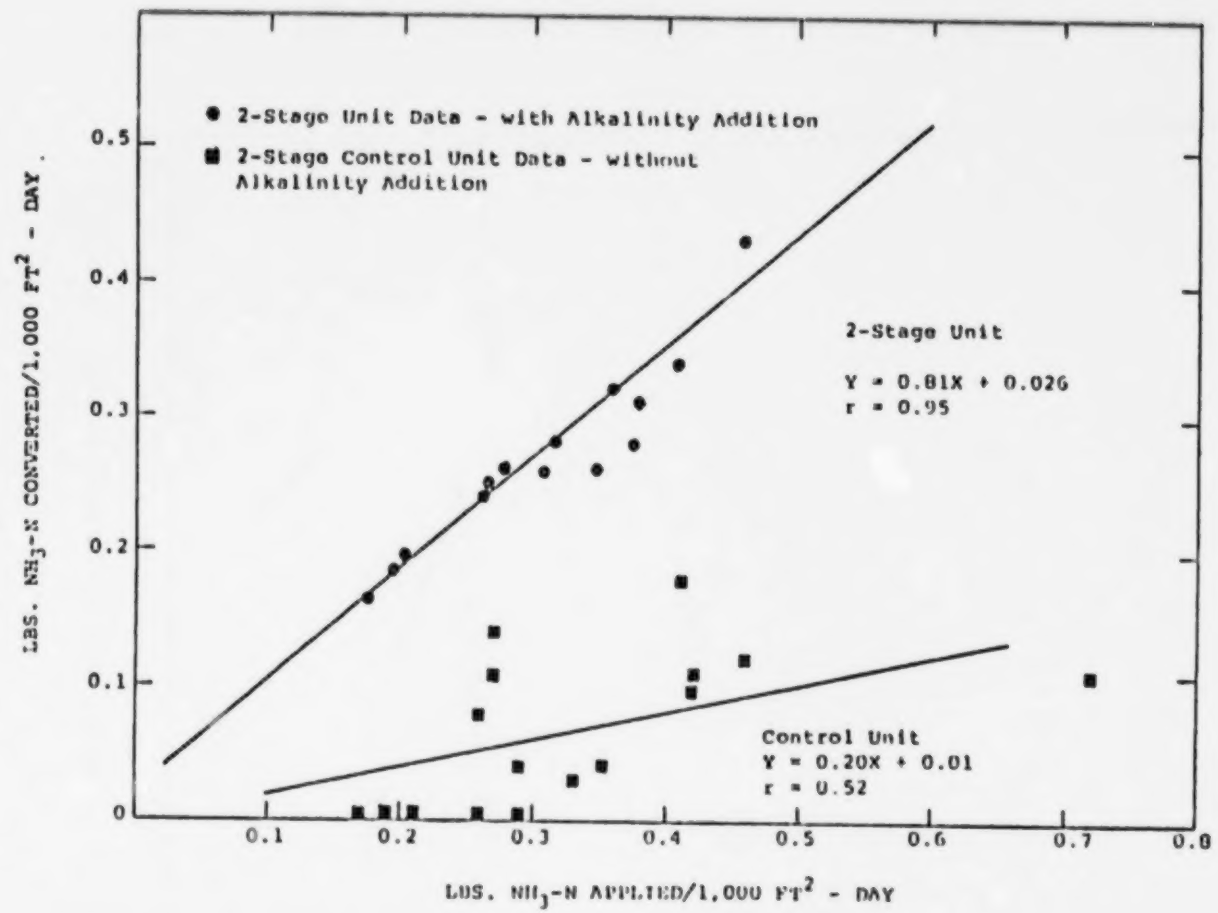

FIGURE 7 : Conversion of Ammonia With and Without Sufficient Alkalinity

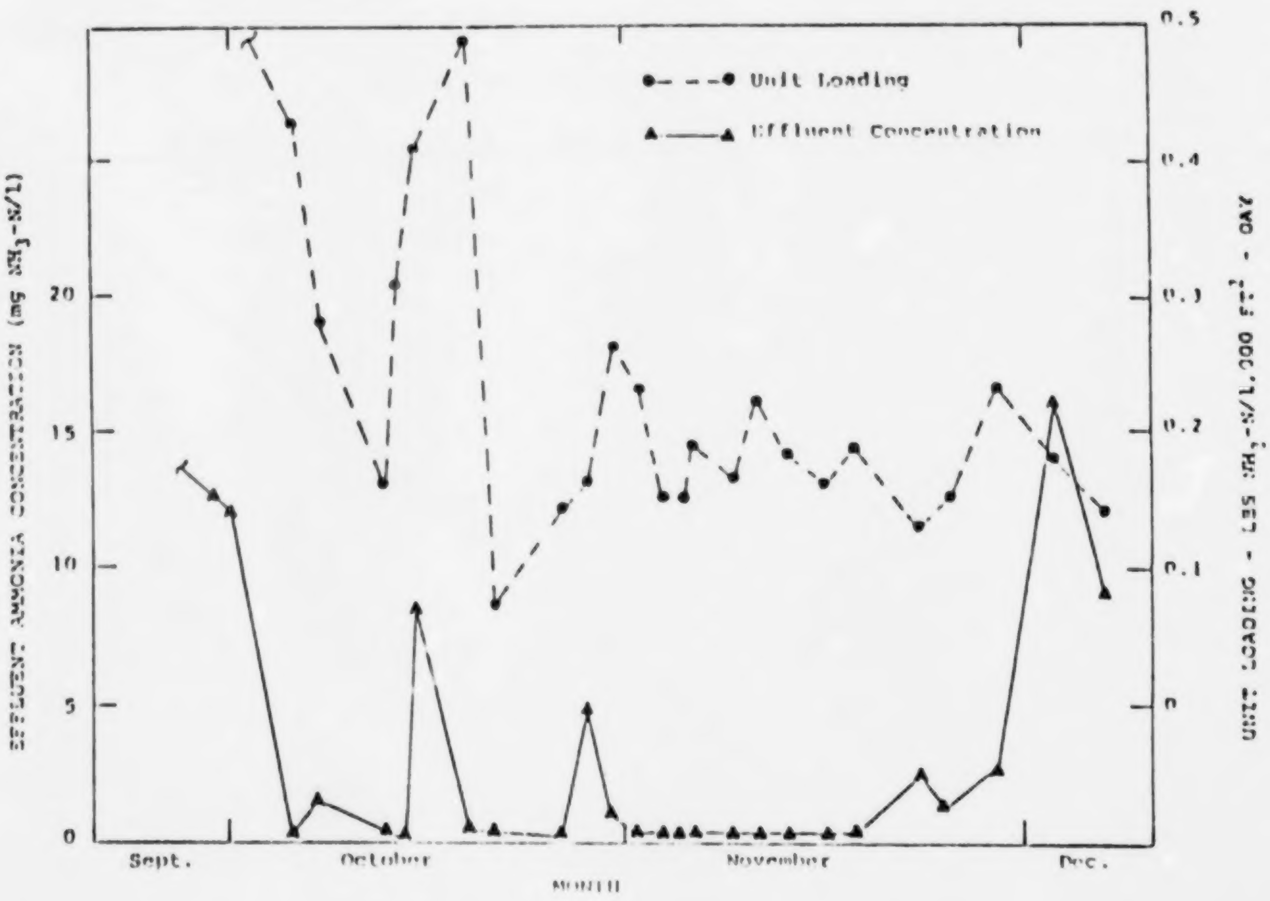

FIGURE 8 : Ammonia Nitrogen Loading and Effluent Concentrations for the Four Stage Pilot Unit

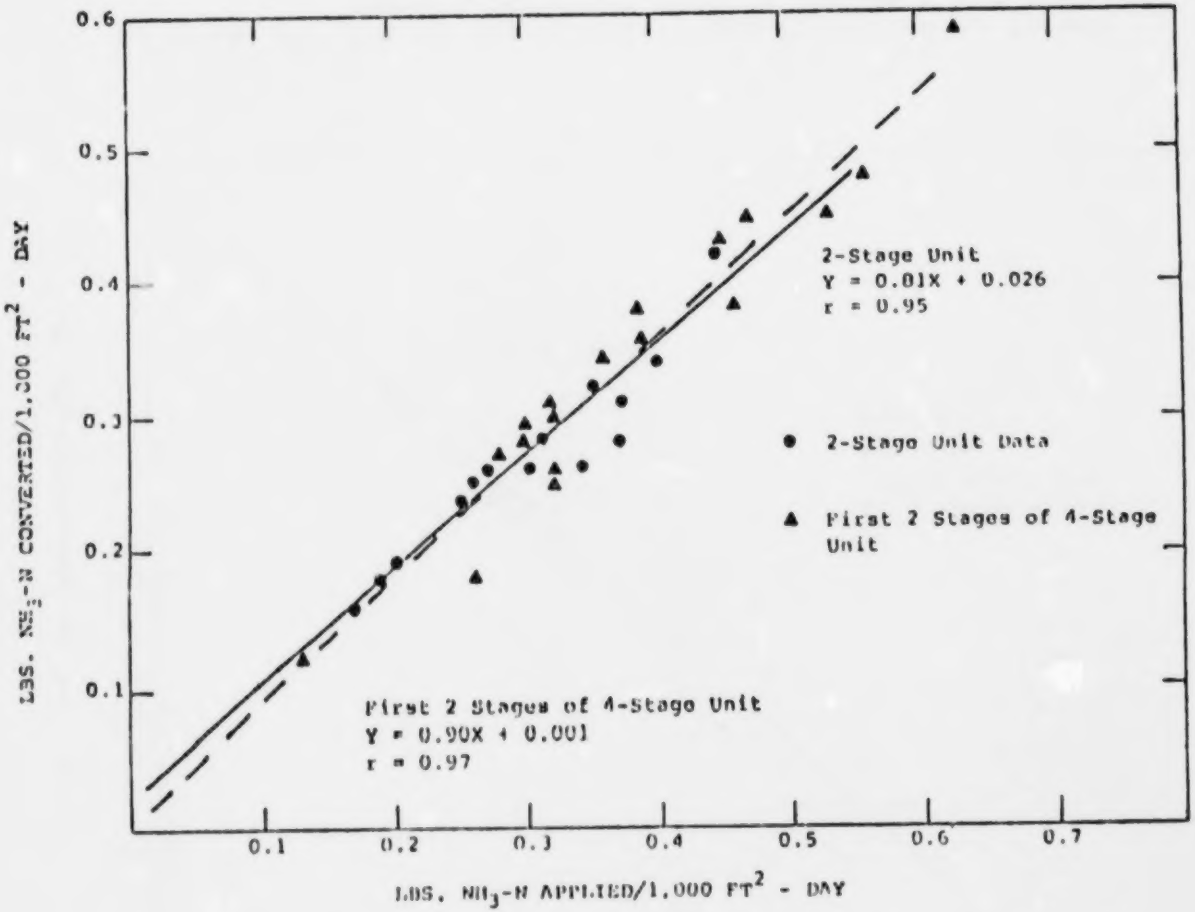

FIGURE 9 : Pounds Ammonia Nitrogen Applied vs Pounds Ammonia Nitrogen Converted Within Two Stages

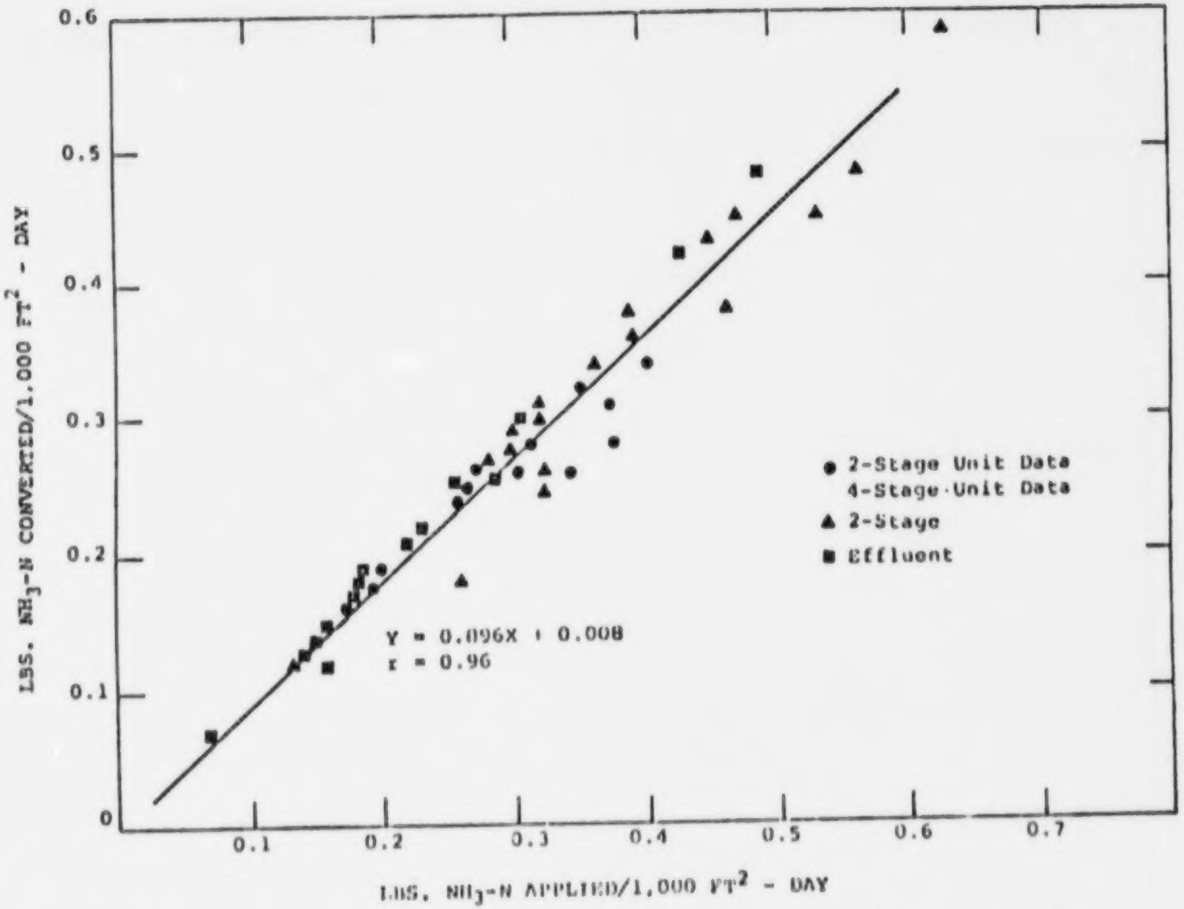

FIGURE 10: Pounds Ammonia Nitrogen Converted vs Pounds Ammonia Nitrogen Applied for Both Two and Four Stage Pilot Units

zed approximately the same surface area, multiple stages appeared to cause increased treatment efficiencies.

However, the two units had different volume to surface ratios. This resulted in the four stage unit having a 33 percent greater hydraulic residence time than the two stage unit. The effect of hydraulic resident time will be discussed in a later section.

SPRING STUDY - LOADING RATES

During April, May and June, 1977, three units were run on the secondary effluent. All three units were four stage units. They were, respectively:

1. EPCO-HORMEL four stage unit - 0.09 gal/ft^2;
2. AUTOTROL four stage unit - 0.12 gal/ft^2;
3. N.U. four stage unit - 0.4 gal/ft^2/

Throughout this study there were several power outages and other mechanical problems both with the units and pumps. As a results, the ammonia loading rates to each unit were more varied with time than occurred in the Fall, 1976 study period. As in both the bench and Fall pilot study, as the unit loading increased rapidly, the ammonia nitrogen concentration in the effluent also increased. However, shortly following the rapid loading increase, the ammonia concentration in the effluent returned to below 1.0 mg NH_3-N/l. During periods of relatively constant ammonia nitrogen loading, the units had effluent concentrations below 1.0 mg NH_3-N/l.

In addition to flow and loading variation, the BOD_5 concentrations in the unit influent also varied throughout the study. The average BOD_5 influent concentration for the Fall study was 10 mg/l, while the Spring study concentrations averaged 22 mg/l. Prior to May 20, 1977, influent BOD_5 concentrations averaged 16 mg/l; and after May 20, the average BOD_5 influent concentration was 29 mg/l. The reason for these different values was a change in the mode of operation at the Marlboro Easterly W.T.P. In the Fall study, the influent averaged 25.5 mg NH_3-N/l, while in the Spring the average was 17.8 mg NH_3-N/l. The increased BOD_5 concentrations with a reduction in NH_3-N concentration caused a reduction in the amount of ammonia nitrogen converted in the first stage of the four stage units. While the Fall pilot and bench scale units converted the majority of ammonia nitrogen in their first stage, the following results were noted during the Spring study.

AVERAGE FIRST-STAGE DATA

Unit	Ratio: gal/ft^2	% NH_3-N Converted	lbs NH_3-N/1000ft^2-day Applied
EPCO-HORMEL	0.09	13	0.76
AUTOTROL	0.12	25	1.10
N.U.	0.4	35	1.29

With an increased BOD_5 loading on all of the units, the conversion of ammonia to nitrate in the first stage was sharply reduced. The unit with the largest hydraulic residence time achieved the greatest percentage of ammonia nitrogen conversion in the first stage, while the unit with the least hydraulic residence time achieved the lowest percent ammonia conversion in the first stage.

The data in Table 2 shows the relationship between influent BOD_5 concentrations and the ammonia nitrogen conversion for various four stage units.

The overall ammonia conversion was good for all units, but the increased BOD_5 loading caused the first stage conversion reaction to either be carried on simultaneously with BOD_5 reduction, or shifted to latter stages within the unit. The net result was reduction of the percentage of nitrification carried out within the first stage. The longer the wastewater remained within the first stage, the greater the percent nitrification achieved.

While the location of active ammonia conversion within the unit shifted to latter stages, overall unit performances remained good. Figure 11 shows the application and conversion rates for ammonia nitrogen in the three four stage units. The N.U. four stage unit achieved the greatest ammonia nitrogen conversion for loadings up to 0.6 lbs NH_3-N/1000ft^2-day. Since the volume to surface ratio of the N.U. unit was approximately four times that of the other units, the wastewater remained in each stage of the N.U. unit approximately four times the other units. The unit with the next highest volume to surface ratio achieved the next highest ammonia nitrogen conversion.

The data indicated that the larger the liquid volume to media surface ratio, the higher the conversion of ammonia nitrogen within the unit. The increased hydraulic residence time within the unit also helped reduce residual BOD_5 within the first stage, allowing more surface area for nitrifying bacteria.

However, all of the suspended solids within the unit must be kept in suspension once they "sluff-off" from the RBC media. If the volume to surface ratio is too large, dead space and associated solids deposition problems may arise. In the N.U. pilot unit, submersible pumps were utilized in each stage to keep the solids in suspension. The scale up factor associated with full scale operations, settling characteristics of the nitrifying sludge, basin geometry, and disc rotational speed will all affect the determination of what is the maximum practical volume to surface ratio that should be used in full scale plant design.

A larger volume to surface ratio will also help reduce the negative effects of sudden peak hydraulic and ammonia nitrogen loadings, The use of staging and optimization of volume to surface ratios are required for optimal unit design.

Nitrate Appearance

The data for nitrate appearance is listed in Table 3. The nitrate analysis was run only during the Fall pilot study, The data indicated that 86 percent of the converted ammonia nitrogen appeared as nitrate nitrogen in the two stage

TABLE 2: BOD$_5$ INFLUENT CONCENTRATIONS AND AMMONIA NITROGEN CONCENTRATIONS BY STAGE

DATE	UNIT	BOD$_5$ (mg/l)	mg NH_3-N/l					
			INF	Stage 1	Stage 2	Stage 3	Stage 4	Effluent
	BENCH							
10/12/76	AUTOTROL	12	25	5.2	1.5	0.5	0.5	0.5
10/25/76	AUTOTROL	6	18	2.8	0.6	0.2	0.2	0.2
5/26/77	AUTOTROL	34	17	13	10	7.0	4.3	4.3
5/26/77	AUTOTROL	64	20	14	10	5.0	2.4	2.4
5/26/77	EPCO-HORMEL	34	17	17	10	6.2	3.8	3.8
6/7/77	EPCO-HORMEL	15	23	16	12	8.2	6.0	5.2
5/26/77	N.U.	34	17	11	2.0	1.6	1.1	0.6
5/31/77	N.U.	64	20	14	6	5	2.7	1.6

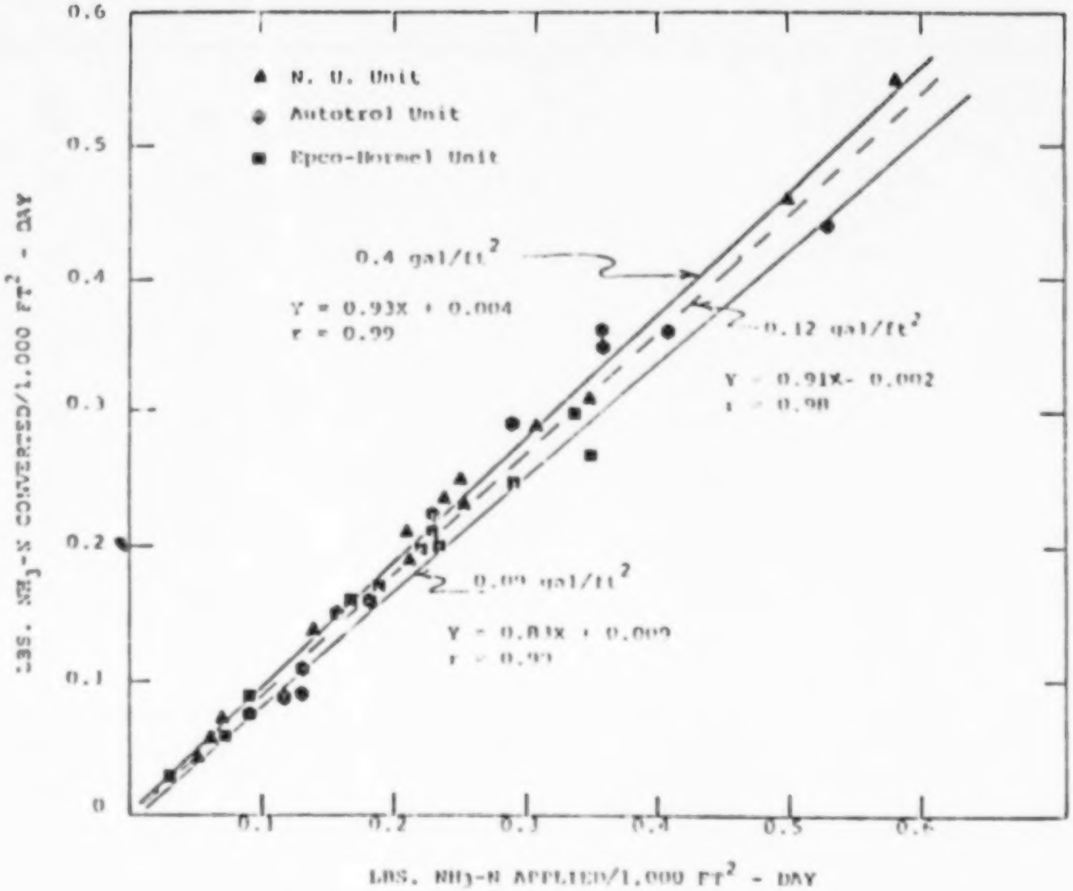

FIGURE 11: Pounds Ammonia Nitrogen Converted vs Pounds Ammonia Nitrogen Applied (Four Stage Pilot Units)

TABLE 3: NITRATE APPEARANCE DATA

DATE	TWO STAGE UNIT			FOUR STAGE UNIT		
	NO_3-N conc (mg/l)		% $\frac{NO_3 \text{ appearance}}{NH_3 \text{ conversion}}$	NO_3-N conc (mg/l)		% $\frac{NO_3 \text{ appearance}}{NH_3 \text{ conversion}}$
	Influent	Effluent		Influent	Effluent	
10/25/77	0.4	16.1	100	0.4	18.4	99
10/27/77	1.2	4.8	100	1.2	16.0	100
11/05/77	0.4	15.8	61	0.4	19.8	72
11/10/77	1.5	21.6	66	1.5	28.5	85
11/12/77	0.6	18.0	82	0.6	19.0	83
11/15/77	0.3	24.0	100	0.3	22.4	86
11/17/77	0.1	21.0	88	0.1	19.2	81
11/22/77	0.3	18.4	72	0.3	20.0	81
11/24/77	0.4	20.5	91	0.4	20.0	88
11/28/77	0.9	25	91	0.9	28.0	96
Average Value			86			88

unit effluent. The four stage unit had an average conversion rate of 88 percent. For the pilot study, 87 percent of the converted ammonia nitrogen appeared as nitrate nitrogen. The bench scale study averaged 93 percent. These results indicate that approximately 90 percent of the converted ammonia nitrogen is associated with nitrate nitrogen in the unit effluent, while about 10 percent is associated with biomass growth requirements.

Solids Production

A summary of the solids data collected during the Spring pilot study is given in Table 4. The values are average values based on the data collected. The N.U. unit had lowest suspended solids concentration in the effluent samples. These samples were taken just as the effluent left the RBC units and represent a mixed sample. When the samples were allowed to settle for one hour, the average suspended solids concentration was approximately the same for all three units (9 mg S.S./l). The N.U. unit had the lowest suspended solids production and lowest suspended solids production to NH_3-N conversion ratio. By comparing the solids production of each unit to the relative hydraulic residence time in each unit, the data indicates that as a unit's volume to surface area increases the net suspended solids production will decrease for a given ammonia nitrogen application rate. This phenomena is analogous to the waste sludge production vs. aeration time in the activated sludge treatment process. The longer detention times allow for solids production to be minimized.

The average total suspended solids concentrations ranged from 16-31 mg S.S. /l for these units. These low concentrations raise questions concerning both the need and effectiveness of a clarifier following RBC nitrification units.

NITRIFICATION: RBC VS ACTIVATED SLUDGE

The bench and pilot studies indicated a high level of nitrification can be achieved using rotating biological contactors. Since the RBC units were treating Marlboro Easterly W.T.P. secondary effluent, a comparison between the RBC effluent and the activated sludge effluent was made. The Marlboro Easterly facility accomplishes nitrification of its secondary effluent with a separate activated sludge system. The ammonia nitrogen concentrations in the final effluents from the Marlboro Easterly W.T.P. were recorded and compared with the RBC effluents.

During the period of the Fall, 1976 pilot study, the activated sludge system had an average NH_3-N effluent concentration of 0.63 mg/l, and the four stage RBC unit had an average NH_3-N concentration of 0.62 mg/l. The Spring, 1977 study period average NH_3-N effluent concentration for the activated sludge system was 0.07 mg/l and the four stage RBC units ranged from 0.61 mg/l to 0.96 mg/l. In this study the RBC units were subjected to variable loading rates. The average NH_3-N effluent concentrations for the RBC units exclude effluent values associated with sharp peaking loads. While the RBC effluent was higher in NH_3-N concentration than was the activated sludge effluent, a good quality effluent was achieved with the RBC units. A more constant loading of the RBC units would have achieved an even higher degree of nitrification

TABLE 4: SOLIDS DATA SUMMARY

UNIT	Effluent (mg S.S./l) mixed	 settled	$\frac{\text{lbs SS produced}}{1000\text{ft}^2\text{-day}}$	$\frac{\text{lbs NH}_3\text{ converted}}{1000\text{ft}^2\text{-day}}$	$\frac{\text{lbs S.S. produced}}{\text{lb NH}_3\text{-N converted}}$
EPCO-HORMEL (0.09 gal/ft^2)	31	9	0.35	0.18	1.9
AUTOTROL (0.12 gal/ft^2)	23	8	0.20	0.22	0.9
N.U. (0.4 gal/ft^2)	16	9	0.04	0.26	0.2

within the units.

SUMMARY AND CONCLUSIONS

Loading rates of up to 0.2 lbs NH_3-N/1000ft^2-day achieved 94 percent ammonia conversion within the unit. Throughout the study periods a constant relationship was observed vetween ammonia nitrogen converted and ammonia nitrogen applied. The data indicated that ammonia nitrogen application rate should be a prime design parameter. Along with ammonia nitrogen loading rate, the volume to surface ratio of the RBC unit also had an effect on the unit's efficiency. A larger volume to surface ratio increased ammonia conversion while causing a decrease in solids production. The increase in unit liquid volume causes an increased hydraulic retention time for the wastewater flowing through the unit. The net effect was similar to those of extending the aeration period in the activated sludge system.

The maximum volume to surface ratio evaluated was 0.4 gal/ft^2. It was necessary to use submerged pumps within each stage to keep the floc in suspension. The problems of scale-up design, basin geometry, and rotational speed were not evaluated.

Future studies on plant scale models should be conducted so that RBC nitrification systems can use optimal design information. Over-design of an activated sludge nitrification system will cause the MLSS to settle poorly. The net results will be a higher suspended solids concentration in the clarifier effluent, and problems maintaining a proper food:microorganism ratio in the aeration tank. Since an RBC nitrification system will not be adversely affected from either of these two problems, it will be advantageous to optimize the volume to surface ratio.

The solids data from this study indicated that a clarifier following the RBC units would be less effective than a filter system. Again, maximizing the RBC volume to surface ratio should increase the length of filter-run, and decrease waste solids handling.

Ammonia nitrogen loading rate and the unit's volume to surface ratio should be used as the major design parameters. A hydraulic loading rate based on gpd/ft^2 is only an indirect approach for constant volume to surface ratios. Use of this hydraulic loading technique will not allow the designer to maximize the unit's efficiencies in terms of ammonia conversion and solids production.

Throughout these studies, 90 percent of the converted ammonia nitrogen appeared as nitrate nitrogen in the unit's effluent. The nitrifying bacteria biomass and other associated losses within the RBC unit accounted for 10 per cent of the converted ammonia nitrogen. Spot checks revealed very low nitrate nitrogen values in the effluent (less than 0.1 mg NO_2/l) and the first stage of each RBC unit was always aerobic.

Since aerobic conditions were maintained within the units, and staging helped with ammonia conversion efficiencies, a minimum of two stages is recommended in RBC design. As a majority of the ammonia nitrogen was converted in the first stage, factors such as loading variation, volume to surface ratio, and BOD_5 loading on the first stage should be evaluated to determine staging requirements.

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PILOT SCALE STUDIES ON THE NITRIFICATION OF PRIMARY AND SECONDARY EFFLUENTS USING ROTATING BIOLOGICAL DISCS AT THE METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

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INTRODUCTION

The Metropolitan Sanitary District of Greater Chicago (District) has been investigating the process of microbial nitrification with different unit operations as a means of removing NH_4-N from its effluents. This report presents and discusses the results of pilot scale studies involving the use of rotating biological disc units for achieving nitrification of primary effluent (single-stage nitrification) and an activated sludge effluent (two-stage nitrification).

Nature of Process

Microbial nitrification is accomplished by a group of chemoautotrophic bacteria. The oxidation of NH_4-N and NO_2-N by nitrifying bacteria may be expressed as:

$$NH_4^+ + \frac{3}{2} O_2 \xrightarrow[\text{ammonia oxidizers}]{} 2H^+ + H_2O + NO_2^- \qquad (1)$$

and

$$NO_2^- + \frac{1}{2} O_2 \xrightarrow[\text{nitrite oxidizers}]{} NO_3^- \qquad (2)$$

Nitrification of sewage can be maintained in aerobic biological treatment processes provided that the following conditions are satisfied:

1. A low enough net growth rate to maintain an adequate population of nitrifying bacteria.

2. Sufficient dissolved oxygen to sustain an adequate rate of nitrification.

3. Satisfactory pH to maintain an adequate rate of nitrification.

4. The absence of substances that inhibit nitrification.

As a system treating a primary effluent, the rotating disc system (RDS) is expected to remove carbonaceous material as well as achieving oxidation of ammonia. Thus, the RDS is expected to provide an environment suitable for the growth and maintenance of organic matter oxidizers and autotrophic nitrifying bacteria.

In a system treating secondary effluent, the RDS system is expected to remove mainly NH_4-N, since the preceeding secondary process will remove considerable amounts of BOD and SS. The RDS system, in this case, would be expected to contain mainly nitrifying bacteria.

The net growth rate of the nitrifying population of an RDS would depend on the NH_4-N loading to the system and the degree of attachment achieved by the biological growth. Generally, fixed growth systems have lower net growth rates than suspended growth systems. Thus, considerably higher solids retention times are possible in fixed growth systems than in suspended growth systems at comparable hydraulic retention times.

Aeration is provided by rotation of the partially submerged discs. Oxygen is absorbed by the thin film of liquid in accordance with Henry's law.

LITERATURE REVIEW

Although the RDS process has been used quite extensively in Europe for more than two decades mainly in small communities for secondary treatment, its developmental work in the U.S.A. began only in 1965. A critical review of this processs and its application to the treatment of municipal and industrial wastes is presented by Antonie (1).

In addition to the utilization of this process for secondary treatment, design criteria also have been developed and reported for nitrification of wastewater based on the data obtained in pilot-scale experiments conducted at various locations (1,2). These criteria have been delineated for both carbon and ammonia oxidation in combined as well as separate wastewater treatment stages. Two important design parameters are the required surface area of the rotating discs and hydraulic retention time to achieve a certain percent of NH_4-N removal from wastewaters. Design curves have been developed relating these parameters to a desired percent removal of NH_4-N contained in sewages of different BOD values. Since the temperature of wastewater influences the growth rate of nitrifying bacteria, temperature correction factors have also been developed to scale down the hydraulic loading rates to ensure nitrification during winter operation (1,2).

The RDS process has been reported to successfully achieve nitrification of a sludge lagoon supernatant liquor, which contained a high concentration of NH_4-N (3). In this study, a pilot-scale rotating biological disc unit (132 liters) was able to oxidize 99.6% of the 918 mg NH_4-N/L contained in the wastewater at a loading rate of 0.2 kg NH_4-N/m^2 of disc surface area/d at 10°C. The ammonia removal averaged 99.8% at 20°C at a loading rate of 0.7 kg NH_4-N/m^2/d at 20°C indicating that at higher temperatures, higher nitrification rates are possible. As a follow-up to this study, successful denitrification of the nitrified sludge lagoon supernatant liquor has also been reported utilizing a submerged rotating biological disc unit (4).

A communication received from Autotrol in February, 1979, one of the companies manufacturing rotating disc systems, indicates that it has provided equipment for 45 installations which are designed for single-stage nitrification and for another 11 which are designed as separate-stage nitrification. These are among a total number of 278 RDS installations which are either under operation, construction, or design (5). The actual number of RDS systems that are operational in the U.S.A. will undoubtedly be more than these numbers, since there are other manufacturers providing these systems.

MATERIALS AND METHODS

Experimental Apparatus

The rotating disc pilot unit was manufactured by the Autotrol Corporation. Figure 1 is a schematic of the pilot unit. The 35 gallon (132 l) semicircular tank was approximately 2 feet (0.61 m) in diameter and 5 feet (1.53 m) in length. The tank was divided into four compartments of equal volume. Lightweight 18-5/8 inch (0.47 m) diameter polyethylene discs were mounted on

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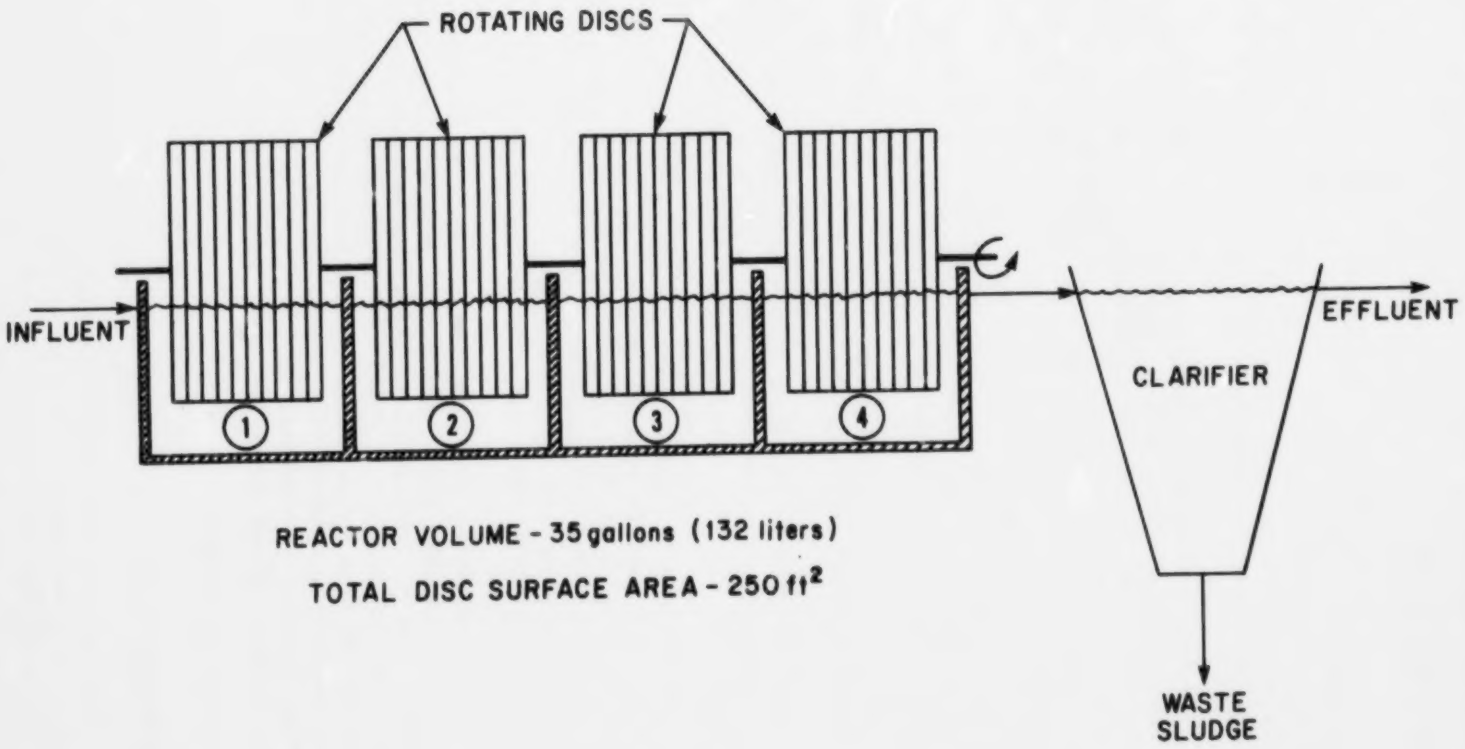
SCHEMATIC DIAGRAM OF THE RDS PILOT PLANT

Figure 1


ROTATING DISCS
INFLUENT
①
②
③
④
CLARIFIER
EFFLUENT
WASTE SLUDGE

REACTOR VOLUME - 35 gallons (132 liters)

TOTAL DISC SURFACE AREA - 250 ft^2

a horizontal shaft which transversed the length of the tank. The discs provided 250 ft^2 (23.2 m^2) of surface area for microbial growth. Approximately 40% of the disc area was submerged in the tank (3). The disc surface area to liquid volume ratio in the pilot unit was 53.4 ft^2/ft^3 (53.4 m^2/m^3). A 1/4 HP electric motor was used to rotate the discs at 7 rpm. A Moyno pump was utilized to control the influent flow to the RDS and, hence, the hydraulic retention time.

For the studies of primary effluent nitrification, a clarifier was used during the latter stages of the testing program (Tests 13 and 14) to determine sludge production. The clarifier was pyramidal in shape. It's size was 6 feet (1.83 m) in height with a surface area of 2.25 ft^2 (0.21 m^2). The volume of the clarifier was 30 gallons (114 l).

Chemical Analysis

In the studies of single-stage nitrification using the RDS process, influent and effluent samples were analyzed for pH, suspended solids, volatile suspended solids, Kjeldahl nitrogen, ammonia, nitrite, nitrate, chemical oxygen demand and biochemical oxygen demand.

In the studies of two-stage nitrification using the RDS process, samples of influent and effluent were analyzed for total Kjeldahl nitrogen, ammonia nitrogen, nitrite, and nitrate.

The chemical analyses used were according to Standard Methods (6) except that modifications were made for the analysis of COD and the nitrogen series (TKN, NH_4-N, NO_2-N, and NO_3-N). These modifications were necessary to adopt these analyses to the automated Technicon procedures used at District laboratories (7).

SOURCES OF SEWAGE

Nitrification of Primary Effluent

In the studies of nitrification of primary effluent, the RDS pilot plant was located at the West-Southwest Plant of the District. As a source of sewage, the RDS pilot plant received primary effluent from primary clarifiers of the Southwest Plant. Characteristics of this primary effluent for the period of July, 1974 to June, 1975 are contained in Table 1. As can be seen, the effluent has sufficient alkalinity to satisfy the needs of the autotrophic nitrifying bacteria in oxidizing the NH_4-N contained in the primary effluent and to prevent any pH changes which will occur because of hydrogen ions produced in the oxidation of ammonia to nitrate. Of course, if a significant drop in pH occurs, it would be unfavorable to the nitrification process. The other constituents present are well below levels considered toxic to carbon and nitrogen oxidizing organisms.

Nitrification of Secondary Effluent

The activated sludge effluent from Battery C of the Calumet Sewage Treatment Plant of the District was used during this phase of the investigation. This conventional activated sludge plant has an influent and effluent with the characteristics shown in Table 2 (average values for the

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TABLE 1

CHARACTERISTIC COMPOSITION OF SOUTHWEST TREATMENT PLANT RAW SEWAGE AND PRIMARY EFFLUENT*

	Raw Sewage	Primary Effluent
BOD	131	88
COD	367	277
NH_3-N	12.68	12.79
pH	7.3	7.5
SS	207	186
TS	832	739
Cd	0.03	0.02
Cr	0.39	0.19
Cu	0.09	0.07
Pb	0.12	0.08
Zn	0.67	0.37
Org-N	11.11	7.92
Hg, µg/l	0.49	0.24
Alkalinity	225	225

All values are in mg/l except pH and Hg.

* Values were averages for the period July 1974 - June 1975.

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TABLE 2

CHARACTERISTIC COMPOSITION OF CALUMET TREATMENT PLANT RAW SEWAGE AND ACTIVATED SLUDGE EFFLUENT**

Parameter	Raw Sewage	Activated Sludge Effluent
Suspended Solids	272 mg/l	27 mg/l
BOD	204 mg/l	25 mg/l
NH_4-N	16.7 mg/l	11.2 mg/l
Org-N	122 mg/l	3.9 mg/l
Total Solids	1034 mg/l	783 mg/l
COD	357 mg/l	104 mg/l
pH	7.3	7.4
Alkalinity	NA*	240 mg/l
Phenols	206 ppb	6 ppb
Hexane Solubles	53 ppb	14 ppb
Cyanide	NA	0.10 mg/l
Cadmium	0.03 mg/l	<0.01 mg/l
Chromium	0.2 mg/l	<0.01 mg/l
Copper	0.5 mg/l	0.13 mg/l
Lead	0.2 mg/l	0.01 mg/l
Mercury	0.3 ppb	0.13 ppb
Zinc	0.6 mg/l	0.02 mg/l

* NA = Not Available

** Values were the averages for the year 1975.

year 1975). The alkalinity is sufficient to meet the requirements of the nitrifying bacteria and to prevent pH drop due to the nitrification process. No constituent is present at levels which could interfere with nitrification.

RESULTS

Primary Effluent Nitrification

During the course of this study, the effect of hydraulic retention time on the degree of nitrification achievable was primarily studied. The hydraulic retention times were able to be manipulated by adjusting the flow rate of the influent to the unit. However, no efforts were made to vary the temperature of the influent. The influent temperatures given in Table 3 were the average temperatures recorded during the indicated period.

A summary of the data obtained at various hydraulic retention times is given in Table 3 with respect to the percent ammonia removals achieved and the concentration of effluent NH_4-N observed.

It is obvious from Table 3 that there was considerable variation in effluent ammonia levels for Tests 6 to 12. The RDS system exhibited wide fluctuations in performance for sewage temperatures of 15°C or less and hydraulic retention times of 2.00 hours or less. Even at a hydraulic retention time of 4.0 hours and a sewage temperature of 27°C (Test 3), effluent ammonia ranged up to 5.4 mg/l, although the average effluent ammonia was 2.3 mg/l.

Effect of HRT

Hydraulic retention time, initially maintained at 5.0 hours at the beginning of the study, was progressively decreased to 1 hours by Test 8. At hydraulic retention times of 2.5 hours and greater, Figure 2, ammonia removals were greater than 80% and effluent ammonia concentrations ranged from 0.6 to 2.5 mg/l. However, ammonia removals fell when the hydraulic retention time was lowered to 1.25 hours. The ammonia removal at this hydraulic retention time averaged in the range of 52 to 78%, and the effluent NH_4-N concentration was in the range of 2.8 to 6.9 mg/l. When the HRT was lowered further to 1 hour (Test 8), ammonia removals were drastically reduced to 16% resulting in an effluent NH_4-N concentration of 13.1 mg/l. It should be noted that the temperature was 9°C during Test 8, which was the lowest during the study. The highest ammonia loading rate of 0.44 also occurred during this test. At a later stage in the study when temperatures were within the range of 15 to 23°C, ammonia removal was considerably higher (Tests 10 and 12) than at 9°C (Test 8) and was about 50%. However, at a temperature of 19°C and a HRT of 1 hour (Test 11), ammonia removal was only 18%. In all the tests conducted at a HRT of 1 hour, only partial nitrification was achieved, and the effluent NH_4-N concentration was in the range of 5 to 13.1 mg/l.

In Test 9, when the hydraulic retention was increased to 2 hours from the 1 hour employed in Test 8, NH_4-N removals increased from 16% to 72%. Test 9 lasted for two months at an average temperature of 13°C. The effluent NH_4-N concentration averaged 3.0 mg/l.

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TABLE 3

EFFECT OF HRT AND TEMPERATURE ON AMMONIA REMOVAL IN
ROTATING BIOLOGICAL DISC SYSTEM TREATING PRIMARY EFFLUENT
JULY, 1974 - JULY, 1975*

Test No.	Period	HRT Hr.	# NH_4-N/ 1000 ft^2/ Day	Temp. °C	Sample	TKN mg/l	NH_4-N mg/l	NO_2-N mg/l	NO_3-N mg/l	% NH_4-N Removal	Range of NH_4-N, mg/l
1	7/74	5.0	.06	20	Influent Effluent	17.4 10.1	11.8 0.6	0.28 0.35	0.19 8.02	95	9.6 - 13.5 0.4 - 1.0
2	8/74	5.0	.07	25	Influent Effluent	22.2 15.3	12.1 1.1	0.08 0.13	0.04 8.65	91	9.9 - 14.9 0.1 - 3.5
3	9/74	4.0	.10	23	Influent Effluent	21.4 7.7	14.5 2.3	0.26 0.39	0.09 10.98	84	5.1 - 20.4 0.5 - 5.4
4	10/74	4.0	.11	18	Influent Effluent	23.2 4.2	15.8 1.0	0.38 0.20	0.48 12.34	94	8.9 - 19.4 0.5 - 2.1
5	11/1/74- 11/17/74	2.5	.19	15	Influent Effluent	23.9 5.8	17.1 2.5	0.36 0.52	0.25 12.94	85	13.5 - 20.8 0.6 - 5.0
6	11/18/74- 12/17/74	1.25	.32	15	Influent Effluent	19.7 9.0	14.4 6.9	0.27 0.84	0.81 9.56	52	9.4 - 17.4 0.1 - 16.4
7	12/18/74- 1/17/75	1.25	.29	12	Influent Effluent	20.2 7.2	12.9 2.8	0.35 0.44	1.38 12.82	78	3.4 - 17.0 0.1 - 10.2
8	1/18/75- 2/21/75	1.00	.44	9	Influent Effluent	22.7 21.7	15.6 13.1	0.29 0.46	1.44 2.68	16	8.9 - 21.0 1.5 - 19.7
9	2/22/75- 4/22/75	2.00	.15	13	Influent Effluent	21.3 12.5	10.9 3.0	0.46 0.65	2.65 9.71	72	2.4 - 15.8 0.6 - 11.0
10	4/23/75- 5/21/75	1.00	.28	15	Influent Effluent	15.9 21.7	9.9 5.2	0.47 1.01	2.70 5.27	47	4.1 - 19.2 1.3 - 13.9
11	5/22/75- 6/22/75	1.00	.32	19	Influent Effluent	17.9 34.2	11.5 9.4	0.47 0.54	0.90 3.78	18	5.3 - 18.1 2.3 - 25.6
12	6/23/75- 7/18/75	1.00	.29	23	Influent Effluent	16.3 32.1	10.3 5.0	0.36 0.39	0.45 5.39	51	3.5 - 18.4 1.2 - 11.2

* Unless otherwise indicated, all the values of the parameters are averages for the duration of the test.

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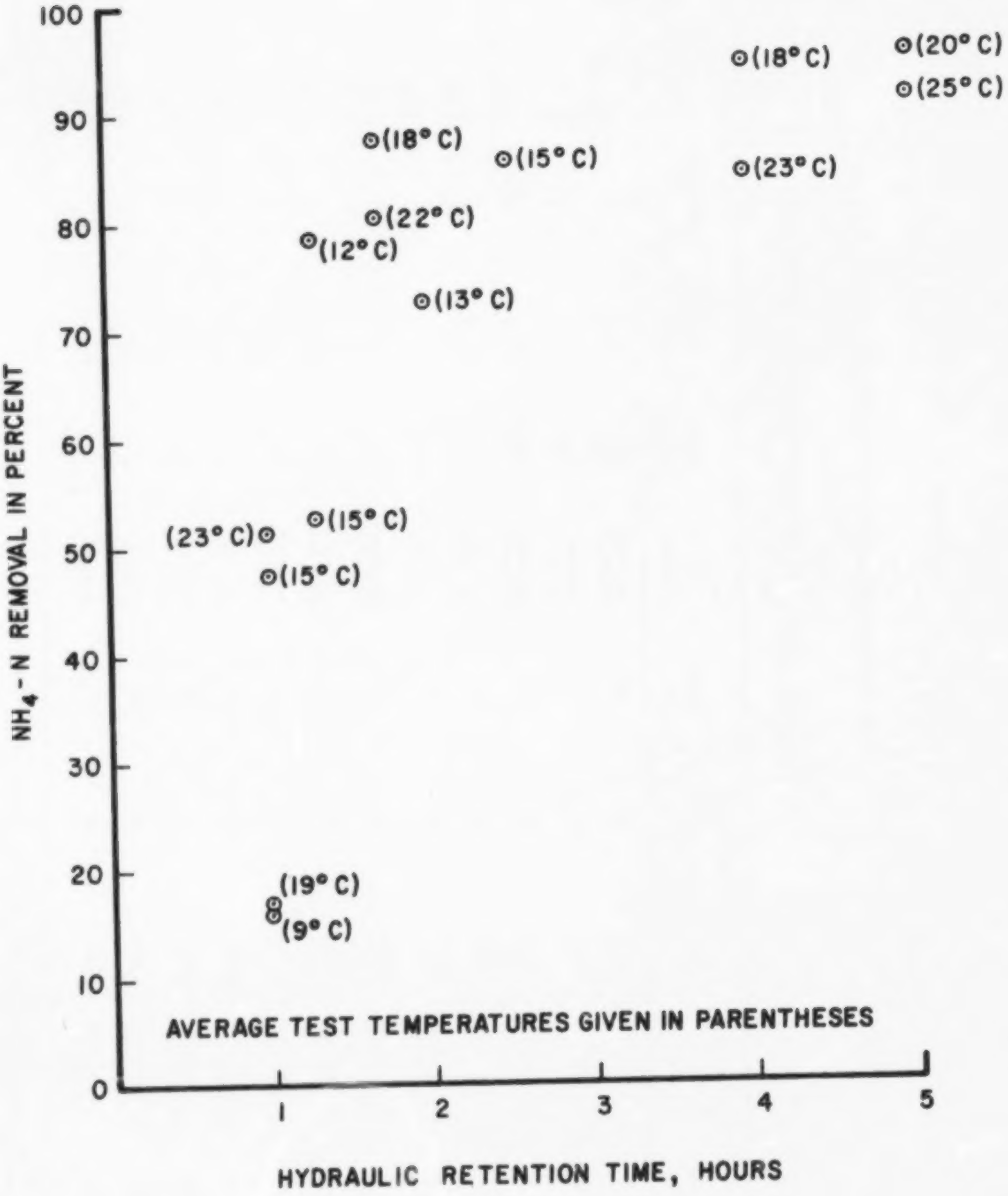
HYDRAULIC RETENTION TIME VERSUS AMMONIA REMOVAL
PRIMARY EFFLUENT NITRIFICATION

Figure 2



The effect of NH_4-N loading rate on NH_4-N removal is presented in Figure 3. This shows that NH_4-N removals were 72% or greater at an NH_4-N loading rate of less than 0.2 lb NH_4-N/1000 ft^2/day. At these loading rates, the effluent NH_4-N concentration was in the range of 0.6 to 3.0 mg/l. Loading rates higher than 0.2 caused only partial nitrification, and the general trend as expected was an increase in the effluent NH_4-N concentration as the loading rate was increased as a result of lowering the hydraulic retention time.

Solids Production and BOD and Suspended Solids Removal in the Nitrifying RDS Unit

In Tests 1 to 12, the feasibility of nitrifying domestic wastewater was studied with no effort being made to clarify the wastewater after it was nitrified. In the following two tests, viz 13 and 14, wastewater was nitrified at temperatures of 22°C and 18°C, respectively, and at a hydraulic retention time of 1.67 hours. In Test 14, small plastic paddles were installed on the perimeter of the rotating discs in each compartment in order to keep the solids in suspension and minimize their accumulation at the bottom of each compartment; where as in Test 13, no such paddles were provided. The results of these two tests are presented in Tables 4 and 5. The results described in Table 4 reflect the overall performance of the RDS in terms of BOD, COD, total solids, suspended solids, NH_4-N removed through the system with and without paddles on the rotating disc unit. With the exception of COD, the removals of suspended solids, BOD, TKN, and NH_4-N were good; the unit with the paddles exhibiting slightly higher removals than the one without them. In Test 14, i.e. the test conducted with the unit containing the paddles, the percent NH_4-N, BOD, and suspended solids removals were 87, 83, and 84, respectively, in comparison to the percent removals of 80, 83, and 77 for the same parameters in Test 13.

The solids production data presented in Table 5 is expressed as suspended solids produced per a) unit of influent suspended solids, b) unit BOD removed, and c) unit COD removed. In the computations, the solids produced is expressed as the sum of the sludge withdrawn from the clarifier and the solids carried over in the effluent.

All three determinations of sludge production show that significantly higher sludge production resulted with paddle operation than without it. The apparently lower sludge production noticed when the RDS unit was operated without paddles may partly be due to the loss of solids accumulated in the unit via biological degradation. Anaerobic decomposition of the accumulated solids was apparent as evidenced by black sludge deposits in the compartments of the RDS. When paddles were installed, such black sludge deposits did not form, and a much grayer sludge resulted than the one encountered in the unit without paddles.

The sludge production data (Table 5) in terms of lbs of suspended solids produced per lb of influent suspended solids, is directly comparable to the data reported by Obayashi *et al* (8) for the District's Southwest Plant. They found this to be the best sludge production parameter (correlation

AMMONIA LOADING VERSUS AMMONIA REMOVAL
PRIMARY EFFLUENT NITRIFICATION

Figure 3

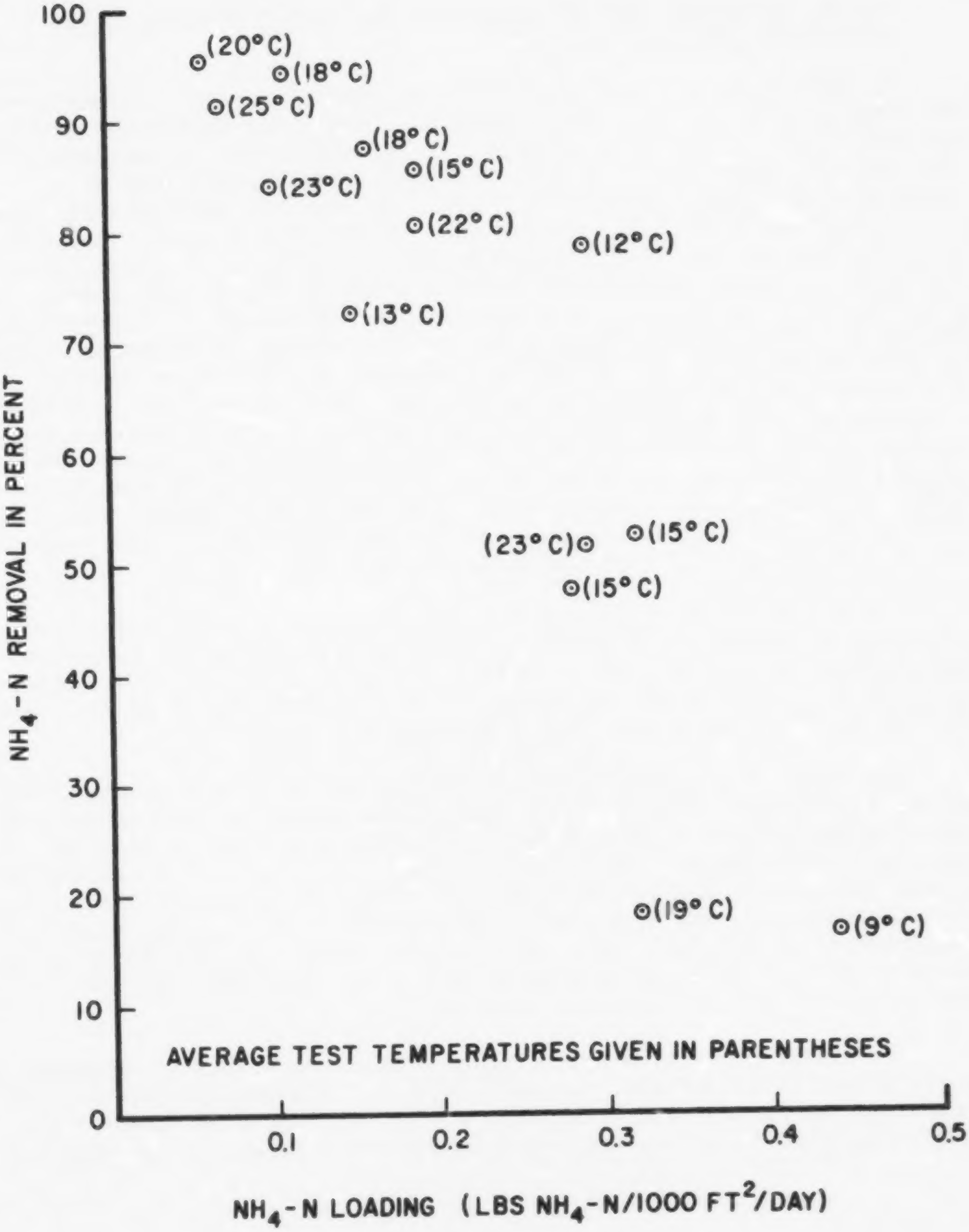


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TABLE 4

PRIMARY EFFLUENT NITRIFICATION
RDS PERFORMANCE, TESTS 13 AND 14
JUNE 30 - OCTOBER 28, 1976*

	Test 13(no paddles) June 30 - August 10	Test 14(w/paddles) Sept. 3 - Oct. 28
HRT, hours	1.67	1.67
Temperature, °C	22	18
NH_4-N Loadings	0.19	0.16
Influent TKN, mg/l	16.9	15.8
Effluent TKN, mg/l	4.2	3.6
TKN reduction, %	75%	77%
Influent NH_4-N, mg/l	11.2	9.5
Effluent NH_4-N, mg/l**	2.2	1.2
NH_4-N reduction, %	80%	87%
Influent NO_2-N, mg/l	0.25	0.37
Effluent NO_2-N, mg/l	0.67	0.61
Influent NO_3-N, mg/l	0.2	0.4
Effluent NO_3-N, mg/l	9.0	5.6
Influent COD, mg/l	277	263
Effluent COD, mg/l	146	121
COD reduction, %	47%	54%
Influent TSS, mg/l	94	96
Effluent TSS, mg/l	22	15
TSS reduction, %	77%	84%
Influent BOD, mg/l	83	90
Effluent BOD, mg/l	14	15
BOD reduction, %	83%	83%
Influent Alkalinity, mg/l	188	194
Effluent Alkalinity, mg/l	132	145
Influent pH	7.4	7.4
Effluent pH	7.3	7.4

* Values are averages for the test.

** Range for Test 13: 0.5 - 10 mg/l, Test 14: 0.0 - 5.6 mg/l.

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TABLE 5

RDS SLUDGE PRODUCTION
JUNE 30 - OCTOBER 28, 1976

	Sludge Production	
	Test 13(no paddles) June 20 - August 10	Test 14(w/paddles Sept. 3 - Oct. 28
$\frac{\text{Suspended Solids Produced}}{\text{Influent Suspended Solids}}$	0.97	1.31
$\frac{\text{Suspended Solids Produced}}{\text{BOD Removed}}$	1.38	1.58
$\frac{\text{Suspended Solids Produced}}{\text{COD Removed}}$	0.70	0.84

coefficient; r^2, of 0.97) in a nitrification study conducted over a four-month period at the Southwest Treatment Plant utilizing an aeration battery receiving a flow of 220 mgd. They reported the sludge production to be in the range of 0.68 to 0.76 lbs per pound of influent suspended solids with an average of 0.72. Clearly, the sludge production from the RDS system is considerably higher than for the single-stage activated sludge system tested by Obayashi et al. It appears that, in this case, the activated sludge system produced less sludge than a corresponding RDS process. One possible reason for the lower sludge production in the activated sludge process is that the battery used in the above study is a newer battery and perhaps is more efficient in terms of oxidizing further the sludge produced than the RDS unit. It is also possible that the activated sludge had a lower F:M ratio than the RDS unit during the respective periods used for the comparison of sludge production.

Secondary Effluent Nitrification

As in the previously discussed study, the only controlled operational parameter was the hydraulic retention time. Other significant variables such as influent ammonia nitrogen concentrations and sewage temperatures were, of course, uncontrolled. The RDS was, therefore, operated at specific HRTs with the objective of determining under what conditions satisfactory ammonia removal could be achieved.

The results of this study were obtained under the following conditions:

1. The activated sludge effluent had ammonia nitrogen concentrations in the range of 8 to 20 mg/l during the period of this study, and averaged 14 mg/l.

2. Sewage temperatures ranged from 9 to 25°C.

3. The effect of hydraulic retention time was tested within the range of 0.6 to 5 hours.

The results of this study are summarized in Tables 6 and 7. Nine different modes of operation were monitored. The operational modes were established at various combinations of hydraulic retention time and temperature.

Table 6 presents the following operational data; hydraulic retention time, ammonia loading, sewage temperature, length of test periods, and effluent ammonia nitrogen concentration. Table 7 shows the following analytical data; influent and effluent total Kjeldahl nitrogen, ammonia nitrogen, nitrite nitrogen, and nitrate nitrogen. Information pertaining to these parameters is given as the average and range for each tested operational mode. Average ammonia removal in percent is also given in Table 7.

The operation of the RDS was begun at a five-hour HRT. At this time, sewage temperatures averaged 25°C. Within 5 days, the oxidation of ammonia to nitrite was established. However, an additional 10 days were required to achieve complete oxidation of nitrite to nitrate.

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TABLE 6

OPERATIONAL PARAMETERS FOR NITRIFICATION OF ACTIVATED SLUDGE EFFLUENT BY PILOT RDS

Test	Hydraulic Retention Time, hr.	NH_3 Loading lb NH_4-N/ 1000 ft^2/day*	Temperature, °C		Days of Operation	Daily Effluent NH_4-N, mg/l		
			Average	Range		Average	Maximum	Minimum
1	5	0.08	25	NA**	11	0.8	1.4	0.2
2	3	0.15	21	18-25	13	1.2	2.3	0.3
3	2	0.22	21	19-23	33	1.3	2.4	0.2
4	1	0.42	12	7-15	54	3.5	9.3	0.8
5	2	0.13	13	11-14	31	1.9	5.6	0.6
6	2	0.18	17	14-19	17	2.2	6.8	0.5
7	1.3	0.21	20	17-23	51	1.2	11.6	0.1
8	0.8	0.48	16	11-18	34	2.3	10.0	0.2
9	0.6	0.83	21	19-22	28	5.1	7.9	0.7

* 1 lb/1000 ft^2/day = 0.205 Kg/1000 m^2/day

** NA = Not Available

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TABLE 7

NITROGEN TRANSFORMATIONS IN NITRIFICATION OF ACTIVATED SLUDGE EFFLUENT BY RDS

Test	Sample	TKN, mg/l		NH_4-N, mg/l		NO_2-N, mg/l		NO_3-N, mg/l		Average NH_4-N Removal, %
		Average	Range	Average	Range	Average	Range	Average	Range	
1	Influent	17.2	14.9-21.5	14.2	10.2-19.0	0.7	0.20-1.50	0.6	0.40- 0.82	94.4
	Effluent	2.3	1.3- 3.4	0.8	0.2- 1.4	6.7	0.50-15.0	4.4	0.34-10.88	
2	Influent	18.6	7.9-22.3	15.8	7.8-18.8	0.5	0.30-0.75	0.8	0.39- 1.88	92.4
	Effluent	2.2	1.3- 3.5	1.2	0.3- 2.3	0.3	0.00-1.00	10.9	5.12-26.88	
3	Influent	18.9	13.8-24.0	15.6	9.0-20.4	1.6	0.65- 3.0	1.5	0.60- 2.40	91.7
	Effluent	2.5	1.7- 4.8	1.3	0.2- 2.4	0.2	0.01-1.67	10.9	6.4 - 16.4	
4	Influent	17.8	9.8-25.0	15.1	6.6-21.2	0.1	0.00-0.42	0.4	0.00- 4.78	76.8
	Effluent	6.6	1.8-25.8	3.5	0.8- 9.3	1.8	0.04-11.00	10.7	1.53-17.35	
5	Influent	13.9	5.7-20.2	9.2	4.0-16.0	0.2	0.01-0.31	1.5	0.00- 4.27	70.3
	Effluent	4.3	1.5-10.2	1.9	0.6- 5.6	0.5	0.07-4.36	11.1	5.00-18.17	
6	Influent	19.1	14.5-25.6	13.0	5.8-15.9	0.1	0.03-0.17	0.1	0.00- 0.63	83.1
	Effluent	5.4	2.2-13.5	2.2	0.5- 6.8	0.3	0.06-0.62	12.9	9.33-15.45	
7	Influent	13.6	6.2-21.9	10.3	3.3-19.5	0.4	0.04-0.89	1.1	0.03- 3.91	88.3
	Effluent	3.4	0.7-14.9	1.2	0.1-11.6	0.1	0.00-0.96	12.2	5.78-16.29	
8	Influent	19.1	9.7-27.1	13.6	7.0-20.0	0.1	0.01-0.42	0.2	0.00- 1.80	83.1
	Effluent	5.2	0.5-14.3	2.3	0.2-10.0	0.4	0.01-1.30	14.7	4.2 - 28.7	
9	Influent	21.8	15.7-27.2	17.2	11.6-23.0	0.1	0.00-0.33	0.1	0.00- 0.6	70.3
	Effluent	7.7	2.1-14.0	5.1	0.7- 7.9	0.7	0.01-2.00	13.1	10.1- 15.6	

Tables 6 and 7 indicate that NH_4-N removals remained relatively constant (88.5% to 94.4%) at HRT values of 5, 3, 2, and 1.3 hours (Tests 1, 2, 3, and 7). The average temperatures at these HRT values were 25, 21, 21, and 20°C, respectively. The average effluent NH_4-N concentration during these tests was less than 1.3 mg/l and ranged between 0.1 and 11.6 mg/l.

When the HRT was decreased to 1 hour in Test 4, the percent ammonia removal decreased to 76.8%. The average effluent ammonia nitrogen concentration over 54 days of operation in Test 4 was 3.5 mg/l and ranged between 0.8 and 9.3 mg/l. This decreased ammonia removal occurred not only because of the reduction of the hydraulic retention time to 1 hour but also because of the low average sewage temperature of 12°C prevailing during this experimental period.

When the HRT was subsequently increased to 2 hours in Test 5, effluent NH_4-N concentrations averaged 1.0 mg/l and ranged between 0.6 and 5.6 mg/l. This effluent NH_4-N concentration resulted at an average sewage temperature of 13°C. Although the average 1.9 mg/l effluent NH_4-N concentration represents a substantial decrease over the 3.5 mg/l NH_4-N concentration of Test 4, the percent ammonia removal increased only slightly from 76.8% to 79.3%. This resulted from sharply depressed influent ammonia levels.

Test 6 represents continued operation at the two-hour HRT but at slightly higher sewage temperatures. The average temperature of Test 6 was 17°C as compared to 13°C in Test 5. Effluent NH_4-N averaged 2.2 mg/l and ranged from 0.5 to 6.8 mg/l, not significantly different from that of Test 5. Percentage ammonia removal increased slightly because of higher influent ammonia concentrations than in Test 5.

At a 0.8-hour HRT and an average temperature of 16°C, effluent ammonia nitrogen averaged 2.3 mg/l and ranged from 0.2 to 10.0 mg/l. Percentage ammonia removal averaged 83.1%. This performance of the unit at 0.8 hours was comparable to that obtained in Test 6 which was conducted at a HRT of 2 hours. As a result of lowering the HRT to 0.8 hours in Test 8, the NH_4-N loading was approximately three times that of the one employed in Test 6.

At a 0.6-hour HRT and an average temperature of 21°C, effluent NH_4-N averaged 5.1 mg/l and ranged from 0.7 to 7.4 mg/l. Ammonia removal averaged 70.3%. This mode of operation yielded the highest effluent NH_4-N concentration and the lowest percentage ammonia removal. Since sewage temperatures were high, averaging 21°C, the decline in effluent quality must be attributed to the short hydraulic retention time and the high ammonia loading.

In spite of the condition of a constant hydraulic flow employed in various test runs, the RDS often exhibited inconsistent performance. This is shown by the effluent ammonia nitrogen data (average and range) given in Table 6. With the exception of Tests 1, 2, and 3, the RDS performance was frequently diminished as indicated by the high effluent NH_4-N concentrations, which are several times larger than the average effluent NH_4-N concentration. At hydraulic retention times of 2.0 hours or less and sewage temperatures of 12 to 20°C (Tests 4 to 8); effluent ammonia levels ranged up to 11.6 mg/l, although average ammonia levels were less than 3.5 mg/l. In Test 6, for a hydraulic retention time of 2 hours and a sewage temperature of 17°C; the average effluent ammonia levels were 2.2 mg/l, but daily values ranged as high as 6.8 mg/l.

From a design criterion standpoint, the relationship between the substrate loading rate (lb NH_4-N/1000 ft^2/day) and ammonia removal is important. Such a relationship is presented in Figure 4 from the data obtained in the pilot scale study. Figure 4 shows that as the ammonia loading is increased, the ammonia removal decreases. This is reasonable, since the finite surface area of the RDS will support a limited amount of biological growth.

Also apparent in Figure 4 is a trend for reduced ammonia removal at low sewage temperatures. This is a result of decreased growth rates of the nitrifying bacteria at low temperatures.

Figure 5 shows the relationship between hydraulic retention time and ammonia removal. As the HRT is reduced, ammonia removal is also reduced. Reduction of the HRT stresses the nitrification process in two ways. Since the HRT is equivalent to the nominal treatment time in the reactor, less time will be available for NH_4-N removal as the HRT is reduced. In addition, reduction of the HRT increases the sewage loading to the RDS. Since the influent ammonia concentration varies over a relatively narrow range, the ammonia loading will generally increase in direct proportion to the sewage flow. Thus, a reduction of the HRT will decrease treatment time and increase the ammonia loading.

DISCUSSION

In general, the results obtained with respect to the hydraulic retention times and hydraulic loading rates required for achieving nitrification in this study of nitrification of both primary effluent and secondary effluent agree with those summarized and reported by Antonie (1). From the results of various pilot plant studies, he found that the nitrification of primary effluent was able to be achieved (90% NH_4-N removal) with a hydraulic loading rate of less than 1.5 to about 3 gpd/ft^2 for primary effluents containing NH_4-N concentration within the range of 40 to 15 mg/l, respectively. These results were obtained above a temperature of 13°C, and the BOD of the incoming wastewaters used in the different studies was within the range of 250 to 80 mg/l. The results reported for the nitrification of secondary effluents at temperatures greater than 13°C and effluent BOD of less than 20 mg/l, indicate that about 90% of the NH_4-N was removed within a hydraulic loading rate of less than 2.0 to 3.5 gpd/ft^2, when these effluents averaged an initial NH_4-N concentration of 40 to 10 mg/l. The higher hydraulic loading rates were possible at lower inlet NH_4-N concentration.

In this study, the HRT for achieving an average NH_4-N removal of 85% or greater from primary effluent was found to be in the range of 1.7 to 5 hours (≈0.5 to 2 gpd/ft^2) for sewage temperatures of 15 to 25°C. In general, decreased HRT and temperature decreased ammonia removal. For example, at an HRT of 1 hour and sewage temperatures of 9 to 23°C, the average percent NH_4-N removal was within the range of 16 to 51%, the higher removal occurring in most instances at higher temperatures. The RDS effluent NH_4 averaged 5 to 13.1 mg/l for individual test runs, while a wide fluctuation was observed on a daily basis within the range of 1.3 to 25.6 mg/l. Of all the runs, only

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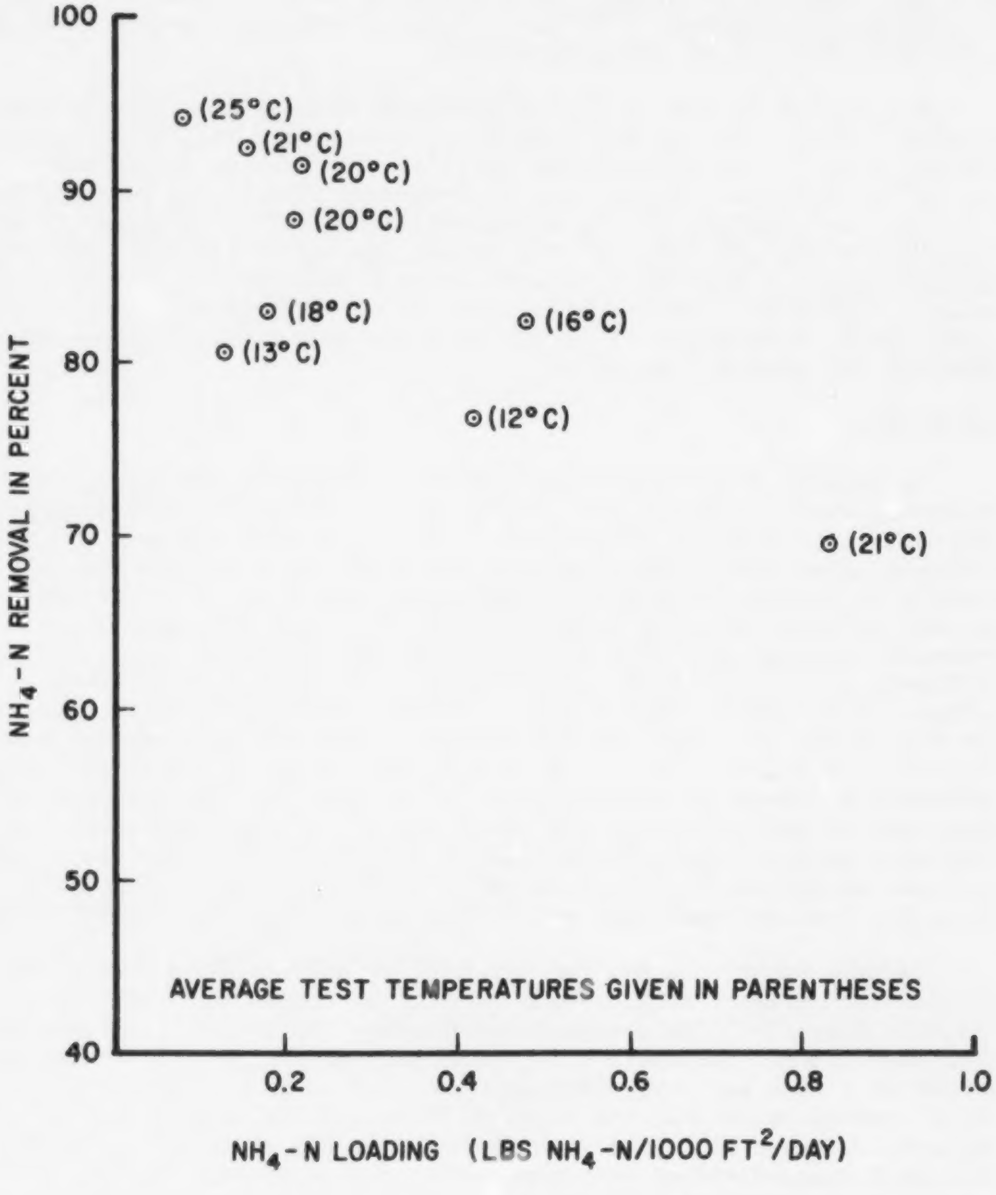
EFFECT OF AMMONIA LOADING ON AMMONIA REMOVAL FOR SECOND STAGE ROTATING DISC SYSTEM
SECONDARY EFFLUENT NITRIFICATION
Figure 4



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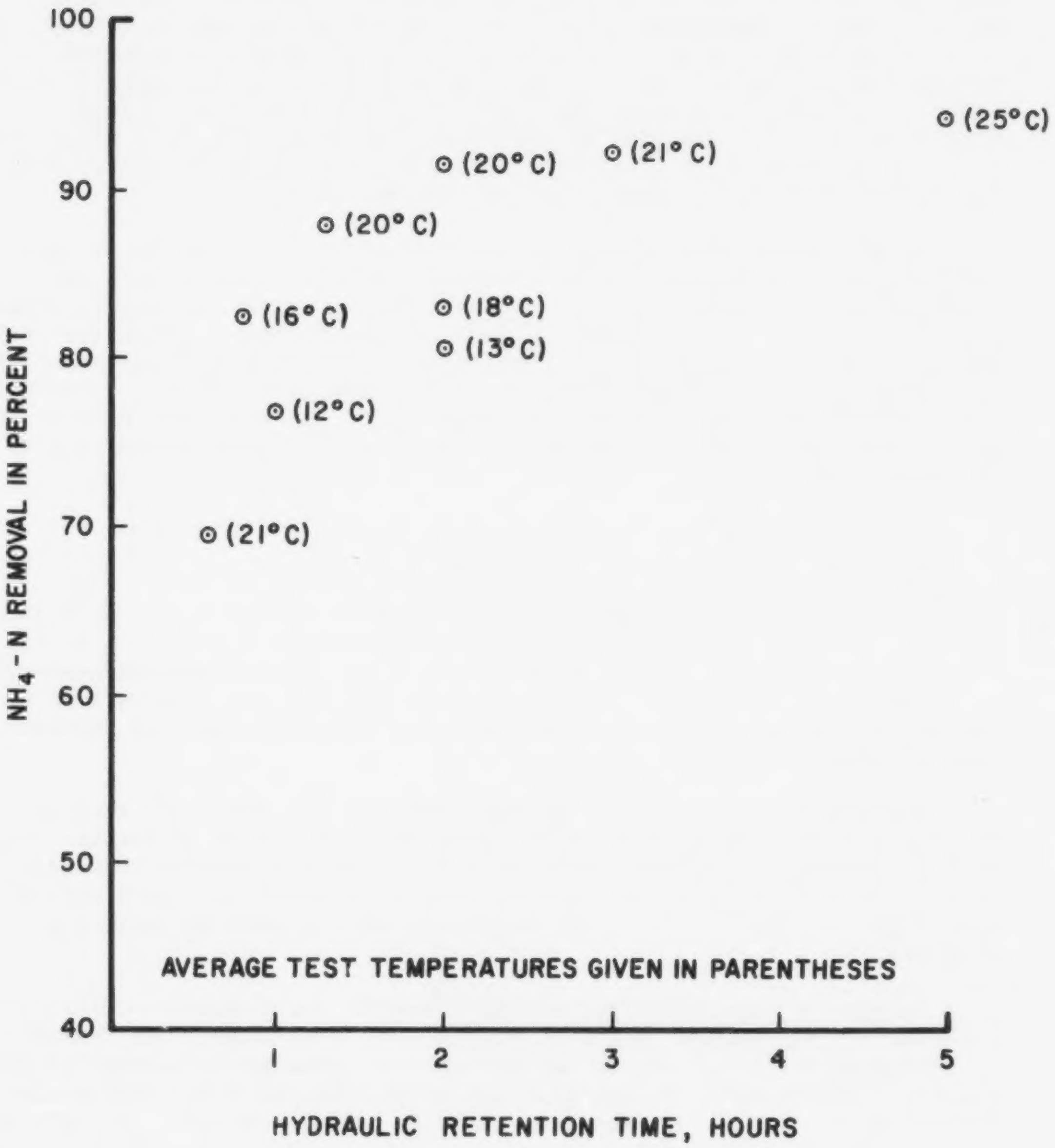
EFFECT OF HYDRAULIC RETENTION TIME ON AMMONIA REMOVAL FOR SECOND STAGE ROTATING DISC SYSTEM

SECONDARY EFFLUENT NITRIFICATION

Figure 5



runs 2 and 4 yielded effluents with the least fluctuations in effluent NH_4-N.

Fluctuations in effluent NH_4-N levels occurred even under the above-noted operating conditions which yielded 85% or greater NH_4-N removals from the primary effluent. Effluent NH_4-N concentration averaged 2.5 mg/l or less, but daily concentration ranged from 0.1 to 10 mg/l.

The degree of nitrification achieved for secondary effluent by the pilot scale RDS unit was comparable to that of primary effluent, in that higher HRTs and temperatures yielded higher NH_4-N removal than lower HRTs and temperatures. The hydraulic loading rates were within the range of 0.7 to 2 gpd/ft^2 for achieving nitrification within an average temperature range of 13 to 25°C for the various runs. However, the wide fluctuations of RDS effluent NH_4-N concentration noted on a daily basis in the case of primary effluent nitrification were also observed in the secondary effluent nitrification. For example, at a hydraulic retention time of 2 hours and a sewage temperature of 17°C, effluent ammonia averaged 2.2 mg/l. However, the levels of effluent ammonia was as high as 6.8 mg/l.

It should be noted that these fluctuations in effluent NH_4-N levels occurred in these RDS pilot-scale studies, despite the fact that HRT remained constant within each of the test runs. Therefore, the fluctuations must be caused by daily changes in influent NH_4-N concentration and/or by the lack of an adequate mass of nitrifiers on the disc surfaces to oxidize the incoming NH_4-N loading within the HRT of the unit. It is, therefore, reasonable to conclude that its performance will be even more erratic when it is subjected to the diurnal fluctuations occurring in sewage flow at a treatment plant.

A recently reported study by Filion, Murphy, and Stephenson (9) alluded to the greater sensitivity of the RDS unit to variation in influent total Kjeldahl nitrogen loading due to fluctuation in flow and HRT. They indicated such performance was not as good as activated sludge nitrification systems. In previously reported nitrification studies performed by the District on activated sludge systems utilizing combined and separate-stage nitrification, the effluent NH_4-N concentration did not fluctuate as widely and was consistently less than 2.5 mg/l once nitrification was initiated and sustained (10,11).

The sludge production in the RDS unit used for the nitrification of primary effluent was considerably higher than the sludge production observed in a plant-scale nitrification battery of the West-Southwest Treatment Plant (1.3 lb vs 0.72 lb of suspended solids/lb of influent suspended solids). This higher production of sludge solids by the RDS unit is obviously undesirable.

In view of the foregoing, although the RDS pilot-scale studies at the W-SW and Calumet Sewage Treatment Plants have demonstrated the feasibility of achieving nitrification of the primary and secondary effluents of these plants, respectively, the installation of the RDS units for full-scale nitrification was not recommended. This is due to the wide fluctuations of effluent NH_4-N concentrations noted on a daily basis and the observed higher sludge production rates with the RDS unit contrary to the stable and lower level of effluent NH_4-N levels and sludge production in the combined and separate-stage activated sludge nitrification systems tested by the District.

CONCLUSIONS

Primary Effluent Nitrification

1. Ammonia removal efficiency of the RDS system correlated directly with hydraulic retention time and temperature, and inversely with ammonia loading rate.

2. At hydraulic retention times of 1.25 to 2 hours and sewage temperatures of 12 to 15°C, the RDS system appeared stressed, producing ammonial removals of 52 to 78% and effluent ammonia levels ranging from 0.1 to 16.4 mg/l. Ammonia loading for these conditions ranged from 0.15 to 0.32 lbs NH_4-N per 1000 ft^2 per day.

3. The study did not yield data on low sewage temperature (10 to 13°C) operation of the RDS consistent with good nitrification (>90% removal). Best low temperature performance was achieved at a temperature of 12°C, a HRT of 1.25 hours, an ammonia loading of approximately 0.3 lbs NH_4-N per 1000 ft^2 - day, and an average ammonia removal of 78%. However, performance was erratic with effluent ammonia concentrations ranging from 0.1 to 10.2 mg/l.

4. Since greater than 90% ammonia removal was only achieved at hydraulic detention times of 4.0 hours or greater, and ammonia loadings of 0.11 lbs NH_4-N/1000 ft^2 - day or less, and only then at temperatures above 18°C, it must be assumed that lower sewage temperature would require at least these levels of HRT and NH_4-N loading combinations in order to produce comparable ammonia removal.

5. Sludge production averaged about 1.50 lbs of suspended solids produced per lb of influent solids.

Secondary Effluent Nitrification

1. Ammonia removal was found to be directly related to sewage temperature, hydraulic retention time, and inversely to ammonia

2. At a hydraulic retention time of 1 hour and a sewage temperature of 12°C, the RDS system appeared stressed, producing ammonia removals averaging 77% and effluent ammonia levels 0.8 to 9.3 mg/l. Ammonia loadings for this condition averaged 0.42 lbs NH_4-N per 1000 ft^2 - day.

3. Hydraulic retention period of 2 hours, ammonia loading of 13 lbs/1000 ft^2 - day, and a sewage temperature of 13°C produced an average effluent of 1.9 mg/l of NH_4-N. However, ammonia removal averaged only 70.3% and effluent ammonia levels fluctuated widely (0.6 to 5.6 mg/l).

4. Ammonia removals exceeding 90% were achieved at hydraulic retention periods of 2 to 5 hours and loadings up to .22 lbs/1000 ft^2 - day

but only with sewage temperature at 21°C or above. In addition, only at these temperatures, hydraulic retention times, and ammonia loadings were effluent ammonia concentrations consistently below 2.5 mg/l.

5. For consistent ammonia removals above 90% at sewage temperatures of 10 to 13°C, hydraulic retention times exceeding 2 hours and an ammonia loading of .22 lb/1000 ft^2 - day or less appear necessary.

Because of the sludge production associated with the use of the RDS system for primary effluent and the fluctuations in effluent ammonia concentrations associated with nitrification of both primary and secondary effluent, the District rejected this system for use in its treatment facilities.

ACKNOWLEDGMENT

The assistance given by Mr. G. Cargill and his associates in maintaining and operating the RDS pilot plant at the Calumet Treatment Plant is greatly appreciated. The authors are also grateful for the analytical support given by the staff of the Quality Control Division of the Research and Development Department.

DISCLAIMER

The results presented herein were obtained by the Research and Development Department of the Metropolitan Sanitary District of Greater Chicago. Publication of this report does not signify that the contents necessarily reflect the views and policies of the District, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

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USE OF ROTATING BIOLOGICAL CONTACTORS FOR NITRIFICATION AT THE CITY OF GUELPH WATER POLLUTION CONTROL PLANT GUELPH, ONTARIO, CANADA

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INTRODUCTION

The City of Guelph is situated on the Speed River about 100 km west of Toronto. A tributary of the Grand River, which empties into Lake Erie, the Speed River had a peak flow of 3450 x 10^6 m^3/d, a low flow of 36.3 x 10^3 m^3/d, and a mean flow of 606 x 10^6 m^3/d in 1978.

Prior to 1979, the Guelph Water Pollution Control Plant consisted of three separate activated sludge type secondary treatment plants, including phosphorus removal, with a total capacity of 46,000 m^3/d. The plant effluent was consistently of good secondary effluent quality at approximately 15 mg/l of BOD_5 and suspended solids and less than 1 mg/l of phosphorus as P. Because of the design loading of the activated sludge system, however, nitrification, that is the oxidation of ammonia to nitrate, did not occur consistently in the plant.

In the summer of 1975, intensive water qualities surveys were undertaken by the Ontario Ministry of the Environment on the Grand River Basin. The data generated by these surveys was used to establish the relationship between dissolved oxygen concentration in the streams and waste loadings from the various water pollution control plants in the Basin. It was concluded that minimum dissolved oxygen concentrations of

4 mg/l would not be realized in the Speed River downstream of the Guelph Water Pollution Control Plant until the nitrogenous oxygen demand of the plant effluent was significantly reduced.

The Ontario Ministry of the Environment proposed the effluent requirements, summarized in Table 1, to be the minimum degree of treatment required in order to meet the dissolved oxygen criteria in the Speed River. For the Guelph effluent, the BOD_{20} to BOD_5 ratio is about 1.5 to 1. This means that the ammonia effluent requirement is 2.0 mg/l.

PROCESS SELECTION

When it became apparent to the City of Guelph that nitrification would become an effluent requirement, studies to determine the organic loading reduction required in the existing plants to achieve nitrification were initiated. While a limited degree of nitrification was achieved, the single stage process lacked stability. It was therefore decided to investigate a two-stage system whereby a separate process would be "added-on" to the existing secondary treatment system.

While there are numerous processes for nitrogen removal, it was determined that for the Guelph application, biological nitrification utilizing an attached growth system would be the least cost solution. Rotating Biological Contactors, or RBC's, were selected as the most economical system for the following reasons:

1. low energy requirement
2. no need for final settling tanks if followed by effluent filtration
3. small land area requirement

PILOT TESTING

In order to obtain design information, a pilot testing program utilizing an RBC was initiated. The City of Guelph procured, installed and operated equipment suitable to carry out the pilot testing as well as performing sample collection and analysis. A pilot scale RBC was leased from the Bio-Systems Division of Autotrol Corporation. It was a 36 disc unit with a total surface area of 23 m^2, media diameter of 0.5 m, a shaft length of 1.32 m and a hydraulic capacity of 0.13 m^3. The tank for the unit was baffled so that there were four separate stages each with an equal number of discs. Influent wastewater for the unit was the effluent from the existing Plant No. 2.

The result of the pilot testing are displayed graphically in Figure No. 1. It shows the specific removal rate of ammonia nitrogen as a function of effluent ammonia concentration. Most of the data was taken at an average wastewater temperature of 18°C. Corresponding removal rates at 16°C and 13°C were then determined using temperature correction factors for nitrification determined by Autotrol Corporation. The specific removal rate follows the classic

Monod relationship with a zero order rate at high ammonia concentrations and a first order rate for low ammonia concentrations with the knee of the curve at approximately 6 mg/l ammonia nitrogen.

Three different rotational speeds were used during the testing period. These corresponded to full-scale peripheral speed, half full-scale speed and full-scale rotational speed. For design purposes, only the half full-scale peripheral speed data was used as the full-scale peripheral speed data was considered to be too optimistic in comparison with other experience and the full-scale rotational speed data was not fully developed due to the length of the testing period.

Diurnal flow conditions existing at the plant were simulated during the pilot testing. For the ranges studied, it appeared that the nitrification process was not upset.

From the data produced during the pilot study, it was determined that at the average wastewater temperature of 16°C, 32 RBC's would be required to achieve the effluent requirement of 2.0 mg/l ammonia nitrogen with an influent concentration of 20 mg/l ammonia nitrogen and a wastewater flow of 55,000 m^3/d.

EQUIPMENT PROCUREMENT

Having established the design criteria for construction of the plant to meet the effluent requirement for nitrification, the City of Guelph advertised public tenders for the "Supply, Delivery, Supervision of Installation, Placing into Operation and Testing of Rotating Biological Contactors". The contract was for the RBC's only and excluded drives and covers.

Specific attention was given in the specifications for the shaft bearings. Greased lubricated spherical roller bearings with spring loaded lip seals and a B-10 life of 200,000 hours were required. Also specified was the inclusion of a hydraulic load cell in the bearing housing biomass thickness on the RBC's.

Recognizing the importance of energy consumption, the tender documents required that the equipment supplier guarantee their equipment's energy consumption for the purposes of both tender evaluation and performance guarantee. Autotrol Corporation guaranteed an energy input of 2.05 kw per RBC.

In addition, a process guarantee was established as shown in Table 2. The guaranteed minimum specific ammonia removal rate was purposely selected at about 10% less than the zero order rate which the pilot study results indicated for 13°C wastewater temperature in order to account for any scale-up factors. The minimum ammonia concentration of 6 mg/l was specified to ensure that the specific removal rate was in the zero order range and thus not ammonia concentration limited. The average influent soluble BOD_5 was selected by the equipment supplier, Autotrol Corporation. The minimum wastewater temperature of 13°C was chosen to correspond to the specific ammonia removal rate. The dissolved oxygen concentration limit of 2.0 mg/l was selected as it is the generally accepted minimum for dissolved oxygen concentration not to be rate limiting for nitrification. The maximum biomass thickness of 3.0 mm was specified in order to limit the stress on the steel central shaft to an allowable level.

During testing, it was required that both the energy consumption and process guarantees be met simultaneously.

Three tenders were received and after lengthy evaluation, a contract was awarded to Autotrol Corporation for 32 air-driven RBC's. The air-drive option was selected for the following reasons:

- simple, infinitely variable speed control system
- considerable capital cost savings over mechanical drive system
- greatly reduced operation and maintenance costs

FACILITY DESIGN

Plant Layout

A plan of the layout of the Nitrification and Filtration Facilities is shown in Figure No. 2. The secondary effluent from Plants No. 1 and 2 was diverted to the Plant No. 3 chlorine contact tank which was converted to a secondary effluent pumping station. All of the secondary effluent is pumped from there to the RBC tanks where the nitrification process occurs. The flow then enters the Automatic Backwash Filters for filtration. Finally, after chlorination and detention in the chlorine contact tank, the flow is measured by a Parshall flume and discharged to the Speed River.

Tankage

To allow the maximum practical degree of staging, the thirty-two RBC's were arranged in four groups of eight in series. The tanks were designed in accordance with the recommendations of the manufacturer, having flat bottoms with the inlet and outlet ends filleted to prevent stagnant areas.

The flow enters the tank through a V-notch weir to provide even distribution across the width of the tank. Precast concrete flow deflectors are installed between adjacent RBC's to prevent short circuiting across the floor of the tank. The direction of rotation of the RBC's is counter-current to the flow direction. Figure No. 3 shows a sectional view of the RBC tanks.

Aluminum baffle plates, with a 100 mm high slot at the bottom are provided between each of the last five adjacent RBC's. The purpose of these baffles is to provide staging downstream of the RBC at which ammonia concentration is less than 6 mg/l. The staging, which in effect creates a completely mixed tank reactor for each RBC, allows maximization of the effluent ammonia from each RBC. Since ammonia nitrogen concentration is rate limiting at values less than 6 mg/l maximizing concentration also maximizes the ammonia removal rate.

Air Drive System

The air drive system consists of three blowers, a discharge air header with a branch to each RBC and individual butterfly valves and diffuser headers for each RBC.

Three centrifugal blowers, each with a capacity of 1.43 m^3/s at a discharge pressure of 20 kPa are provided. To obtain the maximum air drive speed of 1.2 rpm for each RBC requires that two blowers be in operation with a third used as a standby. With only one blower in operation the maximum rotational speed for all RBC's is about 0.9 rpm. Based on the pilot study data, the required process efficiency can likely be obtained in the full-scale installation at a rotational speed of 0.8 rpm.

Each blower is equipped with a motorized butterfly valve for throttling. Placing this valve on the suction side of the blower allows reduced energy consumption when throttling. These valves are operated by a controller which maintains any selected pressure in the discharge header. This feature allows the speed of any individual RBC to be adjusted without affecting the speeds of the other RBC's since the pressure upstream of the control butterfly valve at each RBC is constant.

The diffuser headers are fabricated in thin-wall stainless steel pipe and utilize "Snap-Cap" type coarse bubble diffusers. The headers are removable from the tanks for cleaning or diffuser replacement without removing the fiberglass covers.

Since no part of any RBC is visible during operation, proximity switches are used as speed sensors to detect rotational failure of an RBC.

Covers

The RBC's are protected from weather by one-piece fiberglass covers. Eye-level rectangular ports are provided on one side of the covers. They allow for detection of plugged diffusers by observation of the air cups and manual rotation of the RBC's if required. Removable, press-fit covers allow for bearing maintenance or replacement and access for rotational speed determinations.

The covers are supported on concrete beams which span the width of the tank. Open grating is provided between adjacent beams for a walkway between each RBC.

OPERATIONAL START-UP AND TESTING

Flow Conditions

The RBC's were started up on secondary effluent on 1979-09-19. The flow to the RBC's for the next four weeks was normal flow from the existing secondary treatment plants which included diurnal variations.

In preparation for testing for both process and energy performance, two of the four RBC tanks were shut off and the pumping station was set on manual control to provide a constant flow of secondary effluent to the two remaining RBC tanks, one of which was to be tested. The flow was chosen such that it was slightly less the future design flow of 13,750 m^3/d per tank.

The aluminum baffle plates were placed after the fourth RBC in the tank under test to effectively isolate the first four RBC's. With a surface area of 13,750 m^2 per RBC and a flow of 13,250 m^3/d through the tank, a removal of 6.2 mg/l of ammonia nitrogen was required from the influent to the effluent of the fourth RBC in order to meet the specific removal rate guarantee.

It was assumed that each of the four RBC's under test would operate at the same specific removal rate if their rotational speeds were the same. It would therefore be necessary only to determine the energy input to one RBC for guarantee purposes.

Analytical Equipment

Twenty-four hour composite samplers were installed with sample tubing at the RBC influent and at the slotted opening at the bottom of the baffles at the fourth RBC effluent. The samplers were set to operate in the time proportional mode with one sample being taken each hour. Samples for ammonia determination were preserved with concentrated sulfuric acid added to the sample jar prior to sampling.

The hydraulic load cells and pump were used to determine biomass thickness starting on date of start-up and throughout the testing period.

A temporary orifice plate flowmeter as shown in Figure No. 4,was installed upstream of the butterfly control valve at the third RBC in the tank under test. Pressure and temperature readings taken at the flowmeter allow determination of both ambient air flow and power input to the RBC in accordance with the equations given in Appendix I.

Dissolved oxygen measurements were made at the RBC influent and at each RBC. Wastewater temperature was also determined in the RBC tanks.

Performance Testing Results

Prior to commencement of the performance testing, twenty-four hour composite samples of influent and effluent samples were analysed for ammonia nitrogen to determine whether the system had stabilized at a specific ammonia removal rate in excess of that guaranteed.

On the basis of such testing, the performance tests were commenced on 1979-11-19. It was apparent, however, by 1979-11-21, that the results were not meeting the guarantee requirements and consequently, the testing was stopped.

It was suspected that the reason for the loss of nitrification was excessive carbonaceous oxygen demand loading on the RBC's which inhibited the nitrification population. A simultaneous increase in biomass thickness tended to confirm this suspicion.

The existing secondary plants had been somewhat upset prior to the testing and attempts were made to improve the secondary effluent quality by increasing the aeration tank air supply. At the same time, ammonia removal was monitored using grab samples in order to determine when the testing might resume. More promising results were obtained and testing recommenced on 1979-12-03. The test results are summarized in Table 3.

There was less than 4% difference between average rotational speeds for the four RBC's under test and the maximum difference was less than 10%. From this, it may be assumed that each RBC had the same specific removal rate.

The average ammonia removal across the four RBC's was 6.1 mg/l which is less than 2% below the process requirement of 6.2 mg/l. Given the accuracy of the sampling, analysis, flow measurements and surface area calculations, the RBC's were considered to have met the performance guarantee.

The RBC influent pH averaged 8.0 slightly below the accepted nitrification optimum of 8.5. Typical high influent alkalinity concentrations of 300 mg/l as $CaCO_3$ did not allow a significant pH depression due to the alkalinity consumption of the nitrification reaction.

The average suspended solids and BOD_5 results of 15 and 13 mg/l respectively are typical of the secondary treatment plant performance. The average soluble BOD_5 concentration of 9 mg/l was slightly less than the 10 mg/l maximum limit in the process guarantee.

The average wastewater temperature of $14^{o}C$ was slightly above the minimum required of $13^{o}C$. It is also of interest that there was no change in wastewater temperature from the influent to the effluent of RBC No. 4 in spite of ambient air temperatures about $0^{o}C$ and the uninsulated fiberglass covers.

The RBC influent dissolved oxygen concentration averaged 4.9 mg/l while the RBC No. 4 effluent averaged 5.5 mg/l, an increase of about 12%. This is due in part to the air dirve system and in part to the oxygen transfer capability of the RBC itself. These concentrations far exceed the minimum required of 2.0 mg/l. The influent concentrations are significantly higher than normal secondary effluent because of the aerating effect of the axial flow pumps which lift the secondary effluent to the RBC tanks.

The average energy consumption rate for the tested RBC No. 3 was 1.11 kw. This is almost 46% less than that guaranteed. It should also be noted that this power requirement is at 1.13 rpm. If the process performance can be met at 0.8 rpm as in the pilot study work, the power requirement would be reduced to about 0.5 kw which is about 76% less than that guaranteed.

The biomass thickness on RBC No. 1 in the test tank is plotted as a function of time in Figure No. 5. The data exhibit the classic microorganism growth pattern[1]. Initially, there is a lag phase as the bacteria begin to establish themselves. This is followed by a constant growth phase in which the biomass increases at an expoential rate as all nutrients are in excess. As the nitrifying biomass concentration increases, the ammonia concentration becomes growth limiting and the declining growth phase begins. The system then reaches an equilibrium stage in which the biomass thickness remains relatively constant with minor fluctuations due to variations in influent BOD_5 and ammonia concentrations.

Figure No. 6 illustrates biomass thickness during the test period as a function of RBC tank length. The biomass thickness is linearly related to RBC tank length. This is attributable to the decrease in ammonia concentration through the tank.

Testing was also carried out to determine power input and ambient air flow as a function of rotational speed of the RBC. With the butterfly control valve fully open, the discharge header pressure was varied. The resulting power inputs, ambient air flows and rotational speeds were then recorded. A linear regression was performed on both sets of data expressin power and air flow as power functions of rotational speed. The regression analysis is summarized in Table 4.

For the power input the best correlation is with the third power of rotational speed. This is in agreement with the laws of dynamic similitude for rotating systems[2] in which the following equation is given for the dimensionless group π_3 :

$$\pi_3 = \frac{P}{\rho N^3 D^5} \qquad (1)$$

where $P \equiv$ power
$\rho \equiv$ density
$N \equiv$ rotational speed
$D \equiv$ diameter

In this case ρ and D are constant and

$$P \propto N^3 \qquad (2)$$

The test data points and the linear regression curve are shown in Figure No. 7.

For the ambient air flow the best correlation is with the second power of rotational speed. This can be rationalized by recognizing that the net torque on the RBC is proportional to the ambient air flow in the range of the testing or

$$Q \propto T \qquad (3)$$

where $Q \equiv$ ambient air flow
$T \equiv$ net torque on RBC

It follows that:

$$\because P \propto T \times N \quad \text{(by definition)} \qquad (4)$$

$$\text{and } P \propto N^3 \qquad (2)$$

$$\therefore T \propto N^2 \qquad (5)$$

$$\text{and } Q \propto N^2 \qquad (6)$$

The test data points and linear regression curve are shown in Figure No. 8. Also shown is the comparable data as developed by Autotrol Corporation[3]. The test data is about 10% lower than the Autotrol data. It should be noted that the Autotrol data is for design purposes and is therefore conservative.

Future Testing

The wastewater temperatures are presently (1980-02-06) too low to optimize the system performance. When the temperatures rise again in the summer of 1980, additional testing will be carried to determine the optimum baffle arrangement for staging and the minimum possible rotational speed to minimize energy costs.

COSTS

Energy Costs

The electrical input power at the blowers at two different RBC rotational speeds were determined. With this data the overall efficiency of the air supply system and the energy cost was calculated. These values are summarized in Table 5.

The efficiency of the air supply system decreases significantly at RBC rotational speeds above 0.9 rpm because two blowers must be in operation. The energy cost is reduced by 61% by reducing rotational speed from 1.1 rpm to 0.8 rpm.

Total Annual Cost

The total annual costs are summarized in Table 6. The capital cost is based on the contract prices for equipment and tankage. The electrical energy is based upon operating the RBC's at 1.1 rpm but this may be reduced to 0.8 rpm which will significantly reduce the energy cost. The maintenance costs are estimated at about 1% of the equipment capital cost. The labour cost is based on one man full-time for operation and maintenance.

The cost per unit wastewater flow of 2.7¢/m^3 is about 16% of the total annual cost of a complete secondary treatment plant. It should be noted that these costs do not include for any solids separation after the RBC's. At Guelph, solids separation is accomplished through dual media filtration but only about 15% of the solids loading on the filter is from the nitrification process while about 85% is from the existing secondary treatment system. The pumping cost, which is site specific, is also not included in the above costs.

ACKNOWLEDGEMENT

We gratefully acknowledge the assistance of the City of Guelph's Engineering Department Staff; Mr. R.D. Funnell, P.Eng., City Engineer; Mr. K.J. Bull, P.Eng., Operations Engineer and Mr. J.A. Sanvido, Superintendent of Water and Sewage Works in the preparation of this paper.

APPENDIX I

The following equations were used to determine ambient air flow and power input to the RBC under test.

$$P = 1.011 \, \dot{m} \quad T_2 \left[1 - \frac{p_3}{p_2} \right]^{0.283} \tag{7}$$

where $P \equiv$ power (kw)

$\dot{m} \equiv$ air mass flow rate (kg/s)

$T_2 \equiv$ air temperature six pipe diameters downstream of orifice plate ($^{\mathrm{O}}$K)

$p_3 \equiv$ absolute barometric pressure (kPa)

$p_2 \equiv$ absolute air pressure six pipe diameters downstream of orifice plate (kPa)

$$\dot{m} = \frac{YC_dA_2}{1-\beta^4} \sqrt{2 \ \rho g_c \ \Delta p \times 1000} \tag{8}$$

where $Y \equiv$ compressibility factor (defined below)

$C_d \equiv$ discharge coefficient (determined from tables[4] as a function of Reynolds number at the orifice)

$A_2 \equiv$ orifice area (m^2)

$\beta \equiv$ orifice to pipe diameter ratio

$\rho \equiv$ air density at one pipe diameter upstream of orifice plate (kg/m^3)

$\Delta p \equiv$ differential pressure measured at 1.0 pipe diameters upstream and 0.8 pipe diameters downstream (kPa)

$g_c \equiv$ proportionality factor in Newton's second law ($kg \cdot m/N \cdot s^2$)

$$Y = 1 - (0.41 + 0.35\beta^4) \quad \frac{\Delta p}{p_1} \quad \frac{1}{k} \tag{9}$$

where $p_1 \equiv$ absolute air pressure measured at one pipe diameter upstream of orifice plate (kPa)

$k \equiv$ specific heat capacity ratio for air

$$\rho = \frac{p_1}{RT_1} \tag{10}$$

where $R \equiv$ gas constant for air ($J/kg \cdot {}^{\mathrm{O}}K$)

$T_1 \equiv$ air temperature measured at one pipe diameter upstream of orifice plate ($^{\mathrm{O}}$K)

$$Q = \frac{\dot{m}}{\rho} \frac{T_3 \; p_2}{T_2 \; p_3} \tag{11}$$

where $Q \equiv$ ambient air flow rate (m^3/s)

$T_3 \equiv$ ambient air temperature (^oK)

LIST OF REFERENCES

1. "Unit Processes of Sanitary Engineering", L.G. Rich, J. Wiley and Sons, 1963, pp 9-14.

2. "Essentials of Engineering Fluid Mechanics", R.M. Olson, International Textbook Company, Second Edition, 1966, p 179.

3. "Autotrol Wastewater Treatment Systems Design Manual", Autotrol Corporation, 1979, p C-15.

4. ASME Research Report on "Fluid Meters, Fluid Meter Applications", 6th Edition, p 206.

TABLE 1

EFFLUENT REQUIREMENTS

1. For river temperatures less than 10°C

   BOD_5 - 440 kg/d

   Ammonia - Best Practicable Treatment

   Phosphorus - 1 mg/l

   For river temperatures above 10°C

   Total Oxygen Demand (TOD) * - 1190 kg/d

   Phosphorus - 0.5 mg/l

   $$* \ TOD = BOD_5 \times \frac{BOD_{20}}{BOD_5} + 4.57 \times NH_3 - N$$

2. Effluent loadings from expanded Water Pollution Control Plant to ensure dissolved oxygen in the Speed River downstream of the plant not less than 5 mg/l more than 20% of any day nor less than 4 mg/l at any time.

TABLE 2

PROCESS GUARANTEE CONDITIONS

Specific Ammonia Removal Rate	>	1.5 g/m^2·d
Ammonia Concentration	>	6.0 mg/l
Average Influent Soluble BOD$_5$	<	10 mg/l
Wastewater Temperature	>	13°C
Dissolved Oxygen Concentration	>	2.0 mg/l
Biomass Thickness	<	3.0 mm

TABLE 3

RBC PERFORMANCE TEST RESULTS

DATE	ROTATIONAL SPEED (rpm)				BIOMASS THICKNESS (mm)				RBC INFLUENT							RBC #4 EFFLUENT			RBC #3
	1	2	3	4	1	2	3	4	NH_3-N (mg/l)	pH	SS (mg/l)	BOD_5 (mg/l)	BOD_5(sol) (mg/l)	T (°C)	DO (mg/l)	NH_3-N (mg/l)	T (°C)	DO (mg/l)	P (kw)
1979 12-03	1.20	1.20	1.13	1.15	0.71	0.60	0.47	0.38	18.9	7.9	12	13	10	13	5.2	11.7	13	5.7	1.16
1979 12-04	1.15	1.20	1.13	1.15	-	-	-	-	14.7	7.9	18	11	7	14	4.9	8.8	14	5.8	1.13
1979 12-05	1.13	1.15	1.09	1.13	0.88	.072	0.69	0.60	13.6	8.0	12	14	11	14	5.0	8.3	14	5.3	1.09
1979 12-06	1.13	1.14	1.13	1.13	-	-	-	-	15.1	8.0	14	14	7	15	4.8	8.5	15	5.2	1.04
1979 12-07	1.15	1.13	1.13	1.13	0.88	0.77	.069	.060	17.0	8.3	20	14	8	14	4.7	11.5	14	5.7	1.11
AVE.	1.15	1.16	1.12	1.14	0.82	0.70	0.62	0.53	15.9	8.0	15	13	9	14	4.9	9.8	14	5.5	1.11

TABLE 4

SUMMARY OF LINEAR REGRESSION OF POWER INPUT AND AMBIENT AIR FLOW vs ROTATIONAL SPEED

Assumed Equation	Linear Regression Result	Correletion Coefficient
$P = mN + b$	$P = 1.88N - 1.02$	0.9941
$P = mN^2 + b$	$P = 0.99N^2 - 0.15$	0.9981
$P = mN^3 + b$	$P = 0.70N^3 + 0.13$	0.9988
$P = mN^4 + b$	$P = 0.54N^4 + 0.27$	0.9975
$Q = mN + b$	$Q = 99N - 49$	0.9975
$Q = mN^2 + b$	$Q = 52N^2 - 3.2$	0.9997
$Q = mN^3 + b$	$Q = 37N^3 + 11.6$	0.9989

$P \equiv$ power input (kw)

$Q \equiv$ ambient air flow ($X10^{-3}$ m^3/s)

$N \equiv$ rotational speed (rpm)

TABLE 5

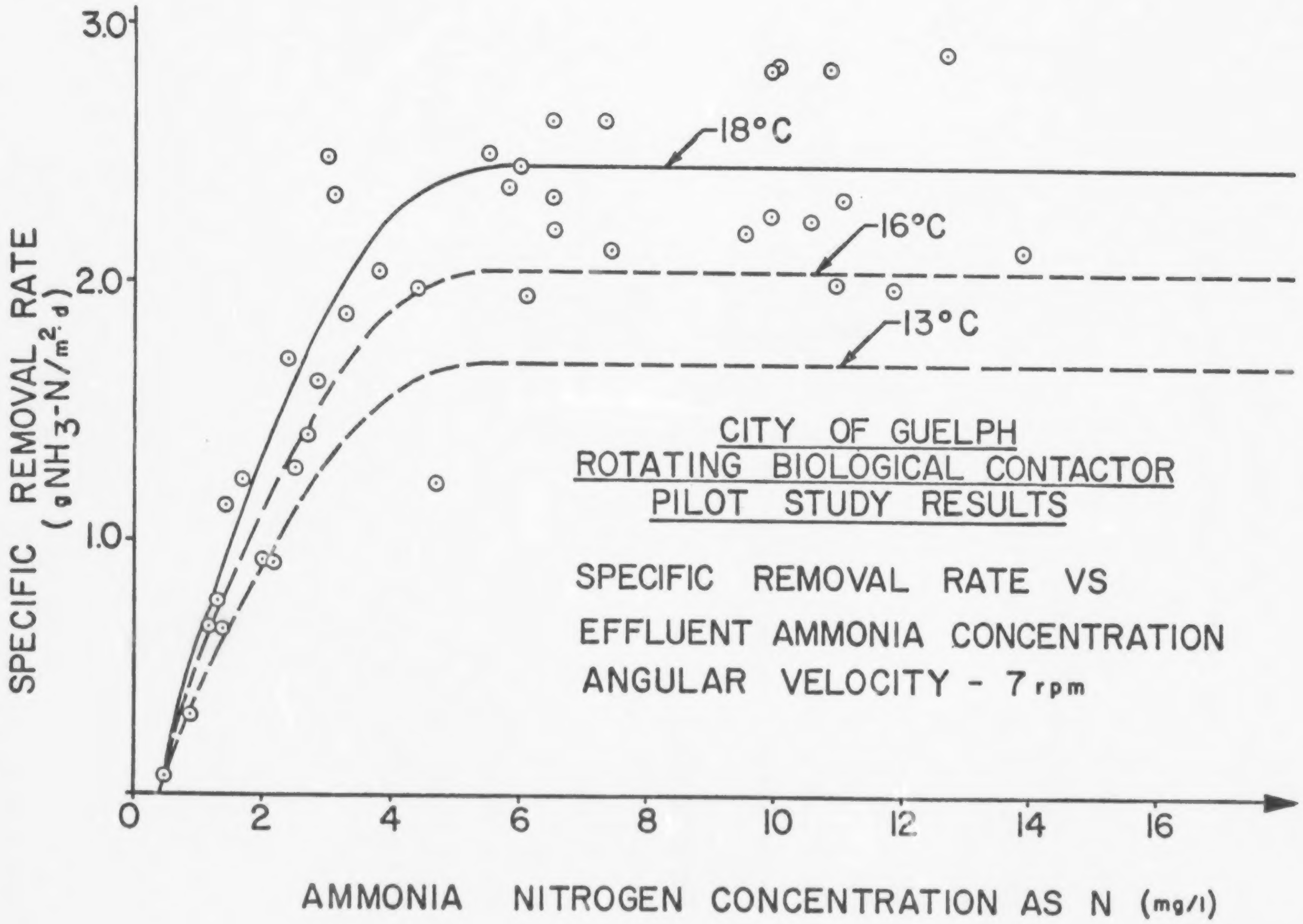
SUMMARY OF ENERGY COSTS

RBC Speed (rpm)	Power Required at each RBC (kw)	Total Power Required at RBC (kw)	Blower Power Required (kw)	Eff. (%)	Energy Cost ($/a) (@ 2.5¢/kwh)
0.8	0.49	15.7	28	56	6130
1.1	1.06	33.9	72	47	15770

TABLE 6

TOTAL ANNUAL COST FOR RBC NITRIFICATION SYSTEM ($Can. 1979)

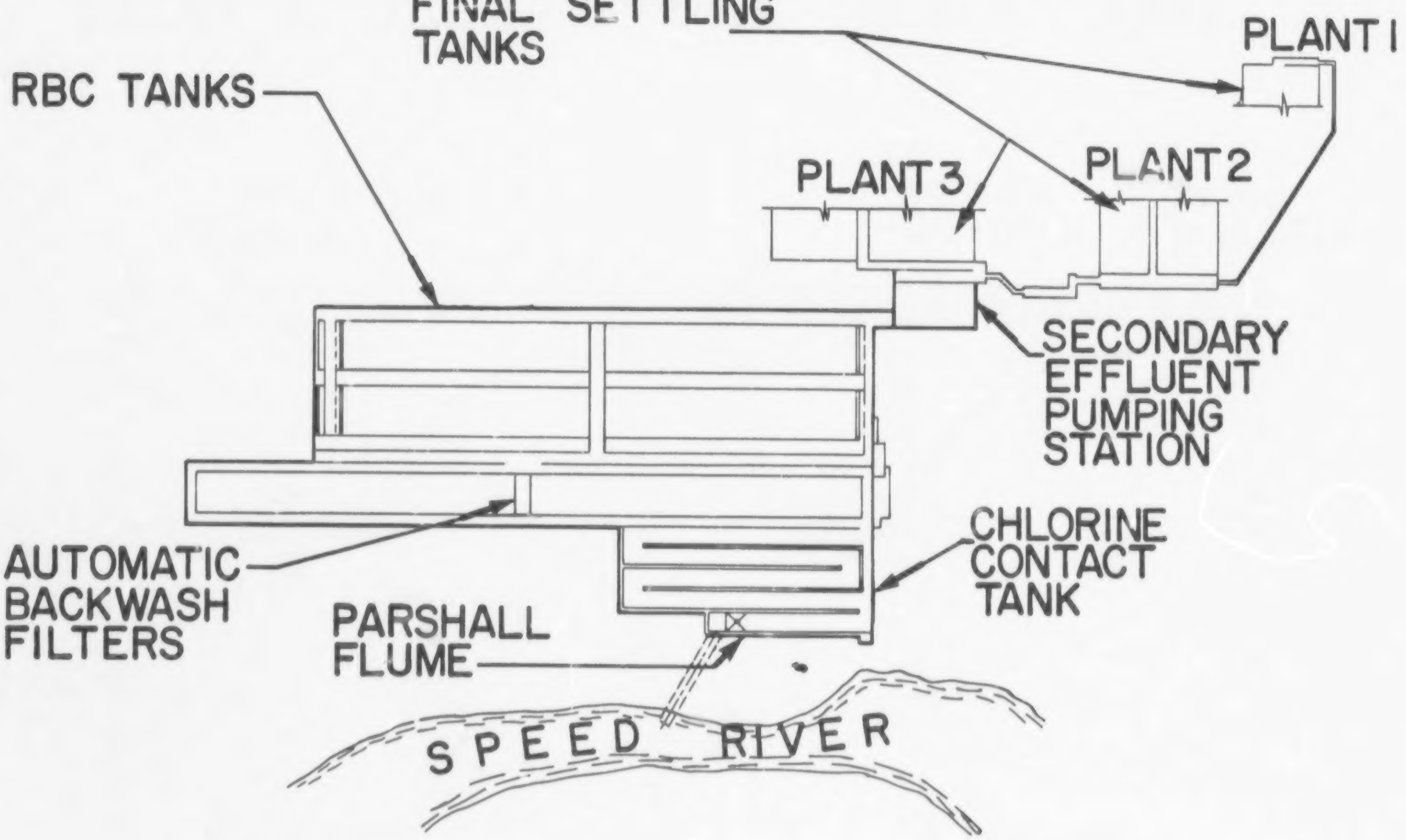
Capital Cost		
Rotating Biological Contactors	1,540,000	
Air Drive System	145,000	
Covers	121,000	
Tankage and Miscellaneous	1,550,000	
Engineering @ 9%	302,000	
Total Capital Cost	3,662,000	
Annual Amortized Capital Cost @ 13%		476,000
Operating Cost		
Electrical Energy	16,000	
Maintenance	17,000	
Labour	30,000	
Annual Operating Cost		63,000
Total Annual Cost		$ 539,000
Cost per unit wastewater flow		2.7¢/m^3



CITY OF GUELPH
ROTATING BIOLOGICAL CONTACTOR
PILOT STUDY RESULTS

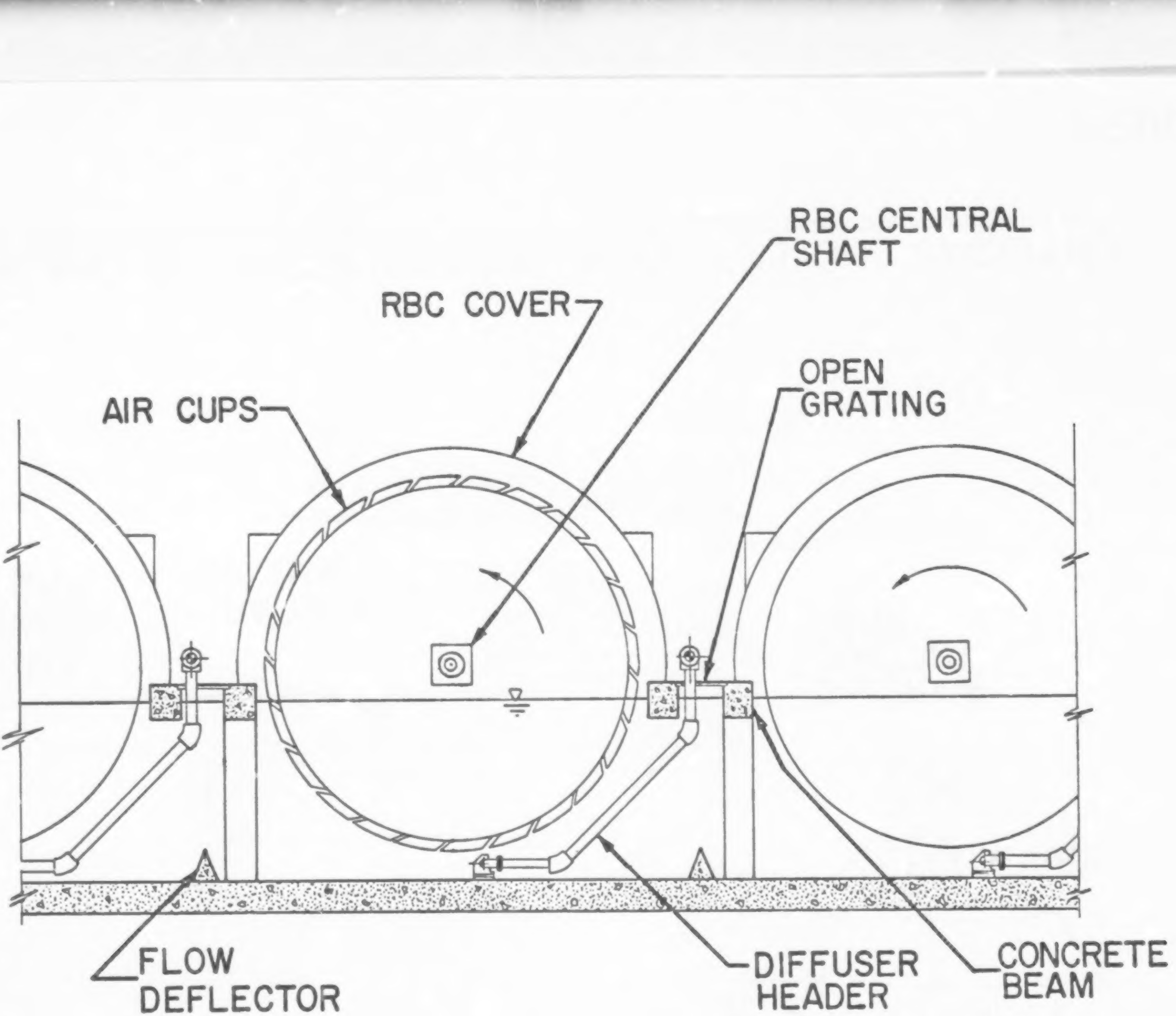
SPECIFIC REMOVAL RATE VS
EFFLUENT AMMONIA CONCENTRATION
ANGULAR VELOCITY - 7 rpm

FIGURE Nº1


FINAL SETTLING TANKS
PLANT 1
RBC TANKS
PLANT 3
PLANT 2
SECONDARY EFFLUENT PUMPING STATION
CHLORINE CONTACT TANK
AUTOMATIC BACKWASH FILTERS
PARSHALL FLUME
SPEED RIVER

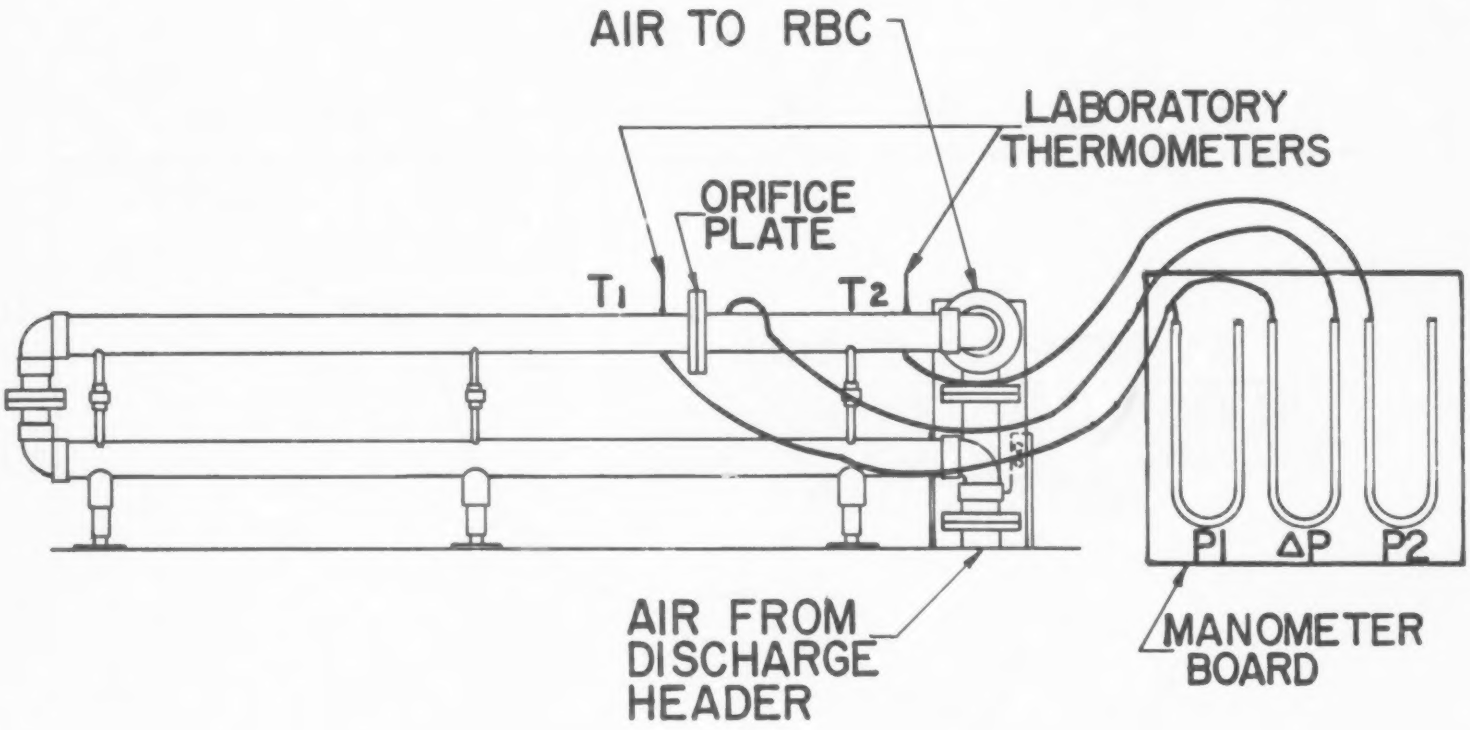
PLAN OF NITRIFICATION & FILTRATION FACILITIES

FIGURE Nº2


RBC CENTRAL SHAFT
RBC COVER
OPEN GRATING
AIR CUPS
FLOW DEFLECTOR
DIFFUSER HEADER
CONCRETE BEAM

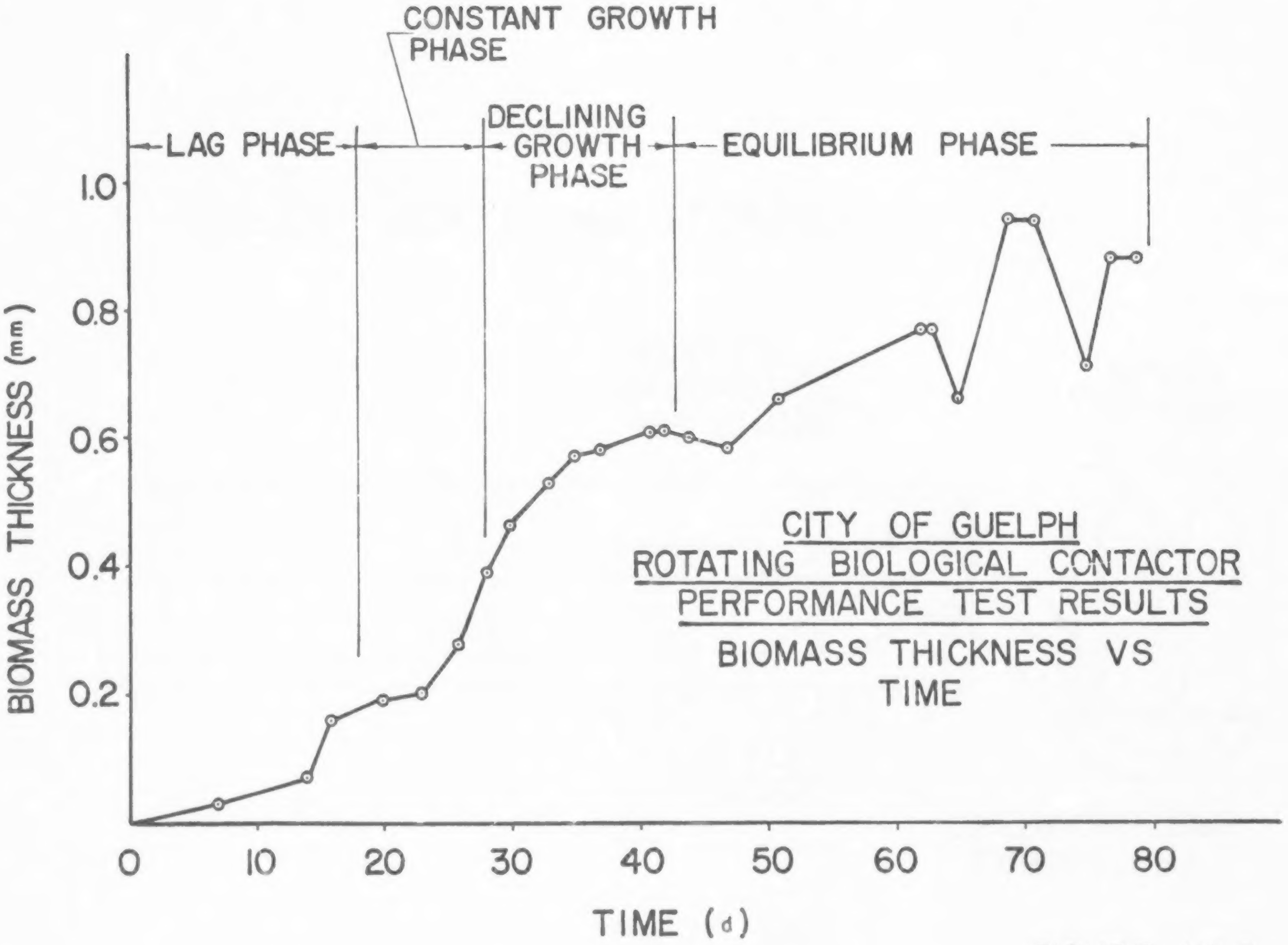
SECTION OF ROTATING BIOLOGICAL CONTACTOR TANKS

FIGURE Nº3


AIR TO RBC
LABORATORY THERMOMETERS
ORIFICE PLATE
T1
T2
P1
ΔP
P2
AIR FROM DISCHARGE HEADER
MANOMETER BOARD

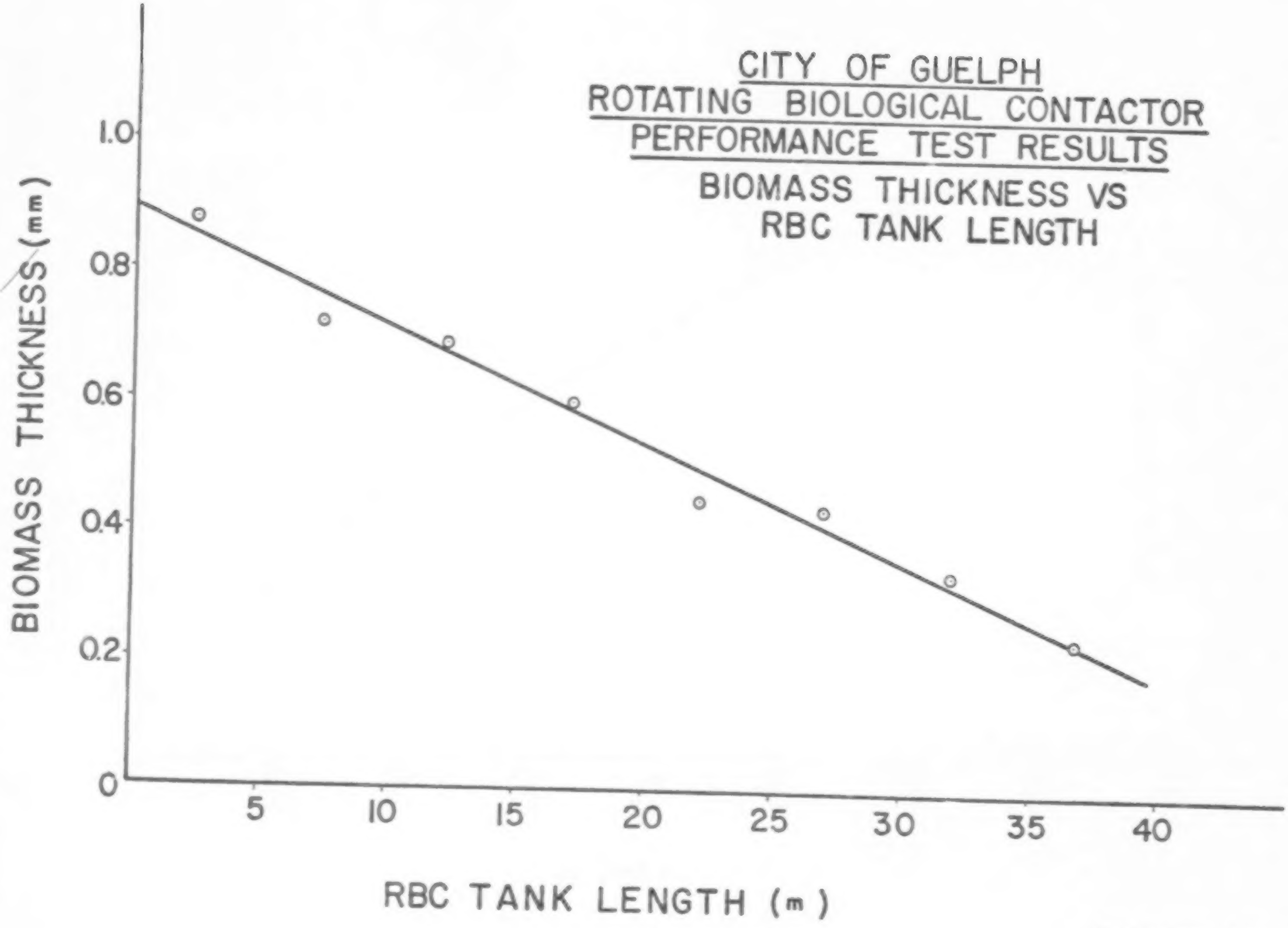
ORIFICE PLATE AIR FLOWMETER

FIGURE Nº4



CITY OF GUELPH
ROTATING BIOLOGICAL CONTACTOR
PERFORMANCE TEST RESULTS
BIOMASS THICKNESS VS
TIME

FIGURE Nº 5



CITY OF GUELPH
ROTATING BIOLOGICAL CONTACTOR
PERFORMANCE TEST RESULTS
BIOMASS THICKNESS VS
RBC TANK LENGTH

FIGURE Nº 6

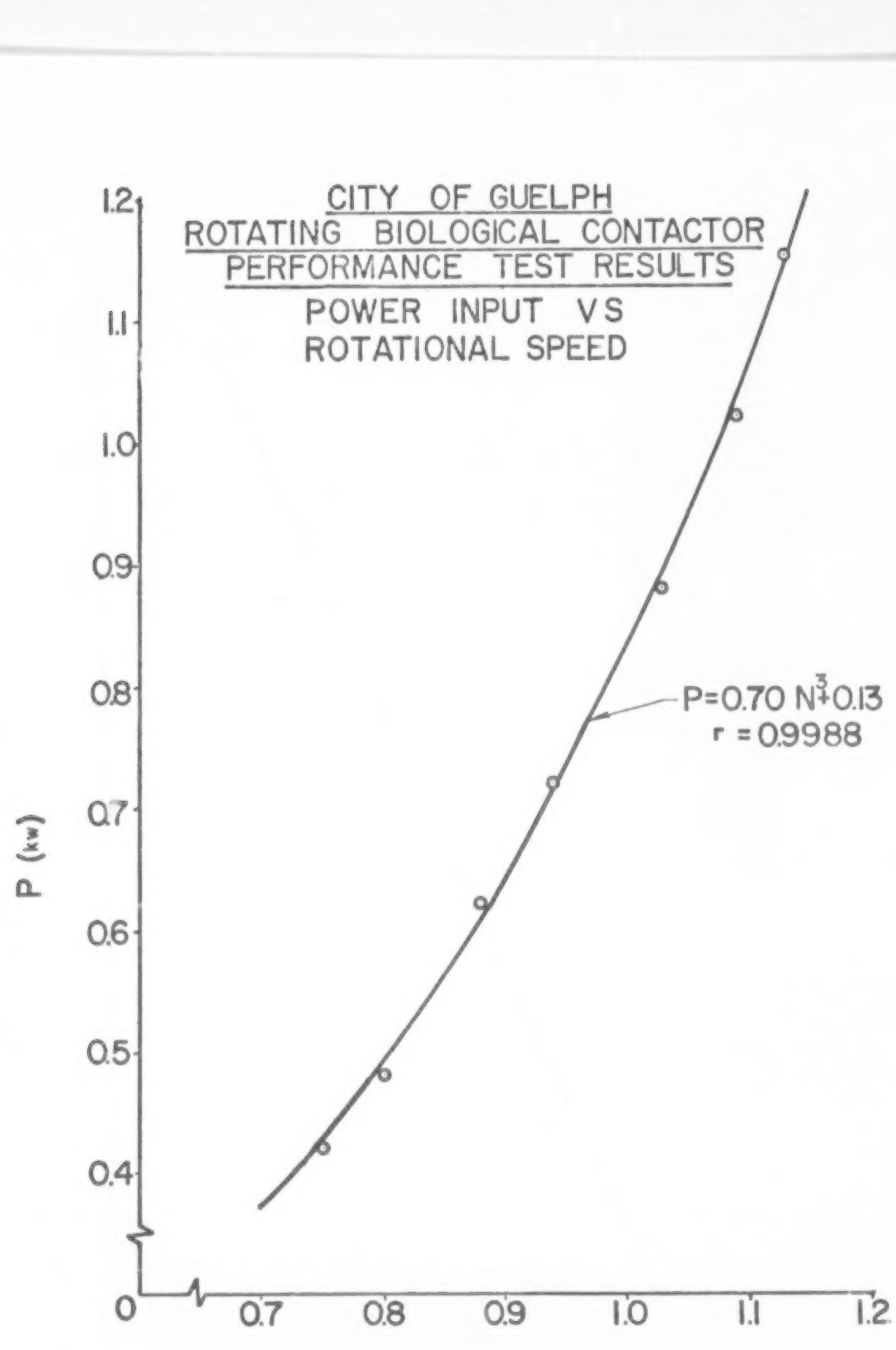
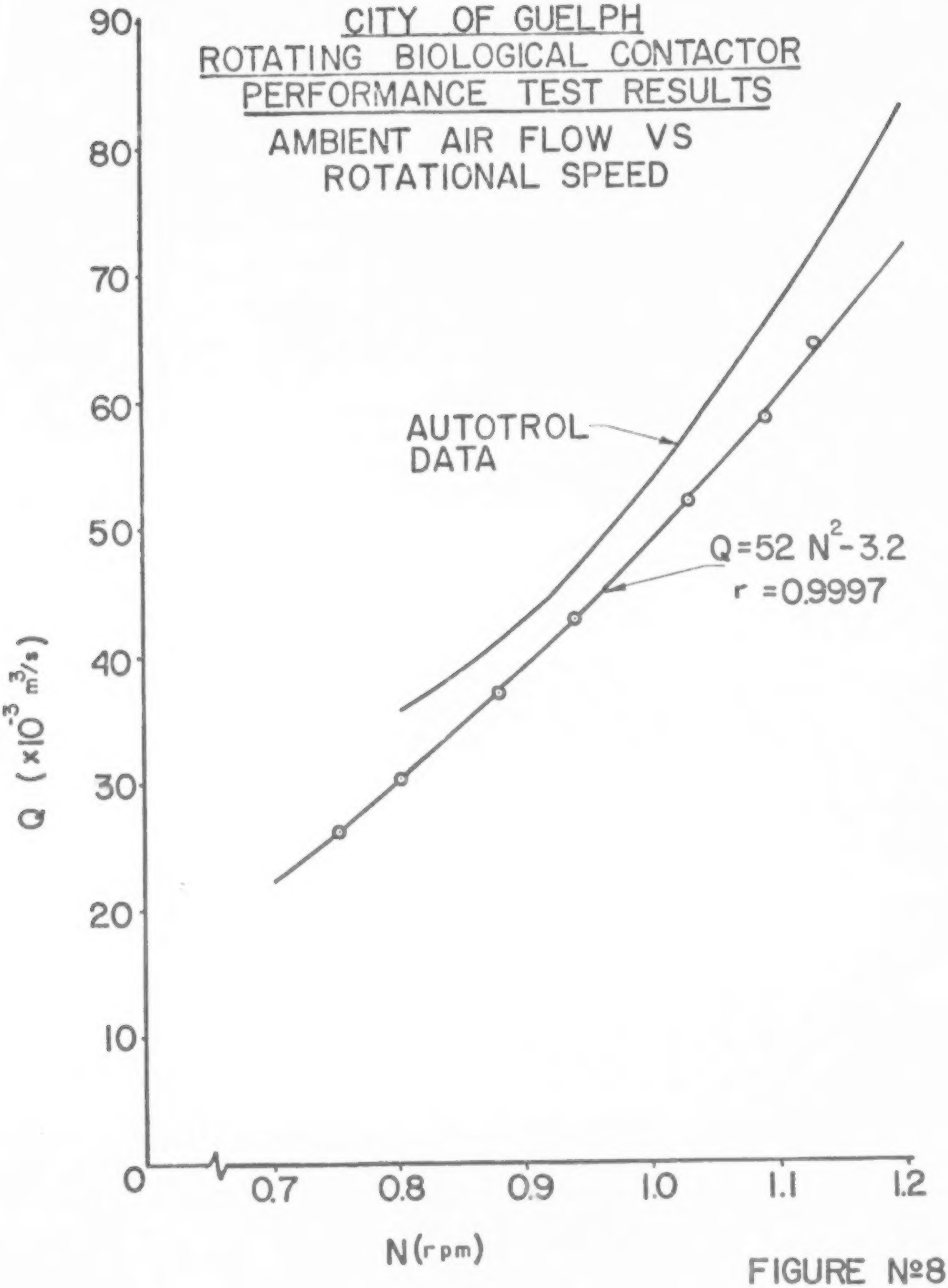

CITY OF GUELPH
ROTATING BIOLOGICAL CONTACTOR
PERFORMANCE TEST RESULTS
POWER INPUT VS
ROTATIONAL SPEED
P=0.70 N³+0.13
r = 0.9988
P (kw)
1.2
1.1
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0
0.7
0.8
0.9
1.0
1.1
1.2
N (rpm)

FIGURE Nº 7



DENITRIFICATION OF NITRIFIED MUNICIPAL WASTEWATER USING ROTATING BIOLOGICAL CONTACTORS

by

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INTRODUCTION

Biological denitrification has been developing rapidly as a means of removing nitrate nitrogen from wastewaters to meet total nitrogen standards when applied for controlling eutrophication in surface waters.

This research utilized rotating biological contactor (RBC) units to perform bench scale and pilot plant denitrification studies of a nitrified domestic wastewater. Tests were performed to determine the effects of hydraulic loading, nitrate concentrations and temperature, and to find the optimum methanol addition requirement.

Specifically, the objectives of this study were: (1) to observe removal efficiencies of nitrate nitrogen under various loading rates; (2) to determine the effect of nitrate concentration within the reactor on removal rates; (3) to determine the effect of temperature on the rate of reaction; and (4) to ascertain the quantity of organic carbon required, as methanol, for optimum removals.

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BACKGROUND

Denitrification is the biological process in which nitrate is partially reduced to gaseous end products and partially converted to cell material.

The conversion of nitrate to cell material is by the assimilatory mode, which follows the stoiciometric equation; (1)

$$14CH_3OH + 3NO_3 + 4H_2CO_3 \longrightarrow 3C_5H_2O_2N + 20H_2O + 3HCO_3$$

where methanol is used for the organic carbon source.

Dissamilatory denitrification follows a series of enzymatic reactions which can be simplified for design purposes as a two-step process. The first step involves the conversion of nitrate to nitrite, and the second step converts the nitrite to nitrogen gas as follows;

$$NO_3 + 0.33CH_3OH = NO_2 + 0.33CO_2 + 0.67H_2O \qquad \text{Step 1 (1,2)}$$

$$NO_2 + 0.5CH_3OH = 0.5N_2 + 0.5CO_2 + 0.5H_2O + OH^- \qquad \text{Step 2}$$

where methanol is used as the organic carbon source. The overall transformation is then represented by;

$$NO_3 + 0.833CH_3OH = 0.5N_2 + 0.833CO_3 = 1.167H_2O + OH^-$$

Since hydroxide ions are produced in this reaction and carbonic acid is present, an increase in bicarbonate alkalinity is expected during denitrification.

Various factors such as; temperature, alkalinity, pH, organic carbon source, and process operating parameters, determine the performance of any denitrification unit.

Dawson, et al (3) performed batch, bench-scale, mixed reactor studies using pure cultures of *Psuedomonas denitrificans* to determine temperature dependencies. The rate of reaction for this test was plotted as a function of temperature and could be represented by an Arrhenius curve. The results of this study are shown in Figure 1.

Alkalinity production for denitrification may be defined stoichiometrically as 3.57 mg of alkalinity as $CaCO_3$ per mg nitrate reduced to nitrogen gas. (1) Good denitrification will take place if the pH is in the 6.0 -- 8.0 range. The optimum pH is in the range of 7.0 to 7.5.

Some idea of proper process operational parameters may be obtained by examination of research involving suspended cell reactors. Moore and Shroeder (4), investigated the effects of cell residence time on cell yield and the rate of dentrification, within an activated sludge system. They concluded that a six day cell residence time was near optimum for low production and thus low methanol requirements. Although greater than nine day cell residence time could reduce these parameters, it would require a substantial unit volume increase for similar flow capacities. The importance of a complete-mix system was noted due to the detrimental effect of short

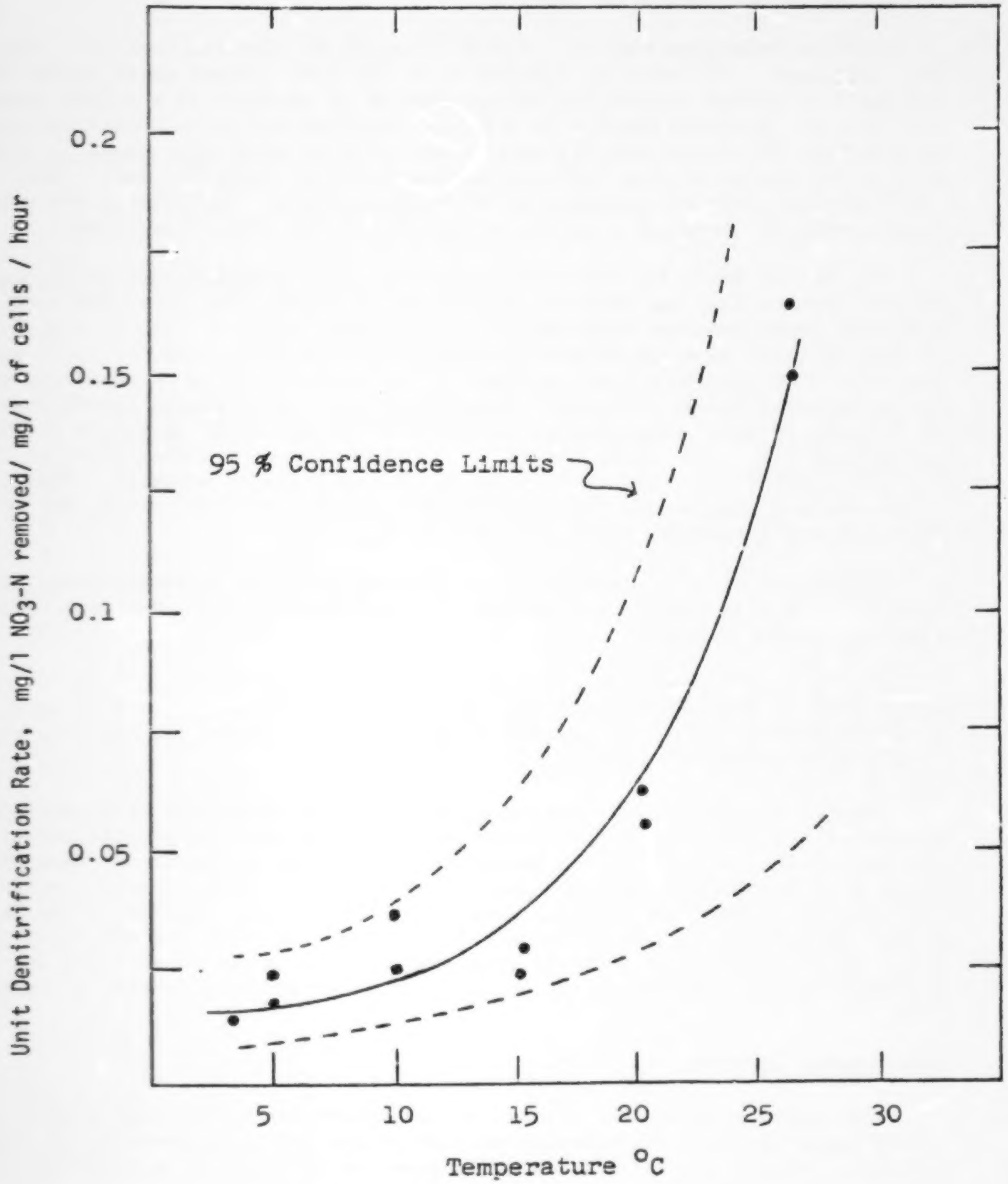


Figure 1 Denitrification Temperature Dependency

R.N. Dawson Data (3)

circuiting on cell residence time.

Various investigations (5,6,7,8,9) using fixed film reactors have also been performed. For example, Duddles et al (5) used a mixed media filter operating as a packed bed denitrification chamber in addition to a solids seperation device. Methanol was fed at a 3.5 mg CH_3OH/mg nitrate nitrogen ratio. Greater than 95 percent denitrification was attained with this system at a flow rate of 2.5 gpm/sf with an influent nitrate level at about 8.7 mg/l. The reaction rate could be determine to be 78.2 mg/sf/min. Backwashing was necessary every 24 hours and required 10 percent of the total flow volume.

The use of RBC's for denitrification has only recently come under investigation. Davies (10) has reported on the use of a partially submerged rotating disc unit (RDU) operated anaerobically for denitrification. The unit provided an area of 11.25 sq m for bacterial growth with a surface area to liquid volume ratio of 789.5 sq m/cu m. All parameters were controlled to be constant, when not being specifically analyzed. Nitrate influent and effluent levels were 60 mg/l and 10 mg/l respectively, the influent being trickling filter effluent spiked with KNO_3. The pH was fixed at 7.0, the temperature maintained at 20°C, and the C:N ratio kept constant at 3:1 by the addition of methanol. Results of the study indicated that the maximum rate of denitrification was 260 mg NO_3-N reduced per square meter per hour.

Murphy, et al (11), reported on a comparison of the denitrification capabilities of a suspended growth system, a submerged RBC unit, and four upflow submerged packed columns.

The suspended growth system operated at a 2.5 hour detention time with an SRT of 3 to 9 days. The RBC unit had a total surface area of 23.2 sq m, was divided into four equal stages and operated at 4 liters per minute giving a 100 minute detention time.

Results of the RBC unit indicate that the denitrification rate was independent of nitrate and nitrite concentrations. The unit did provide efficient and predictable removals at all temperatures normally encountered. Data was reported as removals per unit volume, however, a more useful form for comparison is on a surface area basis. From the reaction rate temperature dependency curve, removals at 5°C and 20°C could be determine to be 93 and 435 mg/sq m/hour, respectively. The effluent from this unit was reported to quite low in suspended solids, indicating the possibility of eliminating further clarification.

EXPERIMENTAL EQUIPMENT

This paper presents the results of laboratory bench and pilot plant research investigations which evaluated some of the design parameters involved in denitrifying a nitrified municipal wastewater effluent using rotating biological contactors.

Initial laboratory bench scale experiments were conducted at Northeastern University's Environmental Engineering Laboratories over a period of 12 weeks in the spring of 1976.

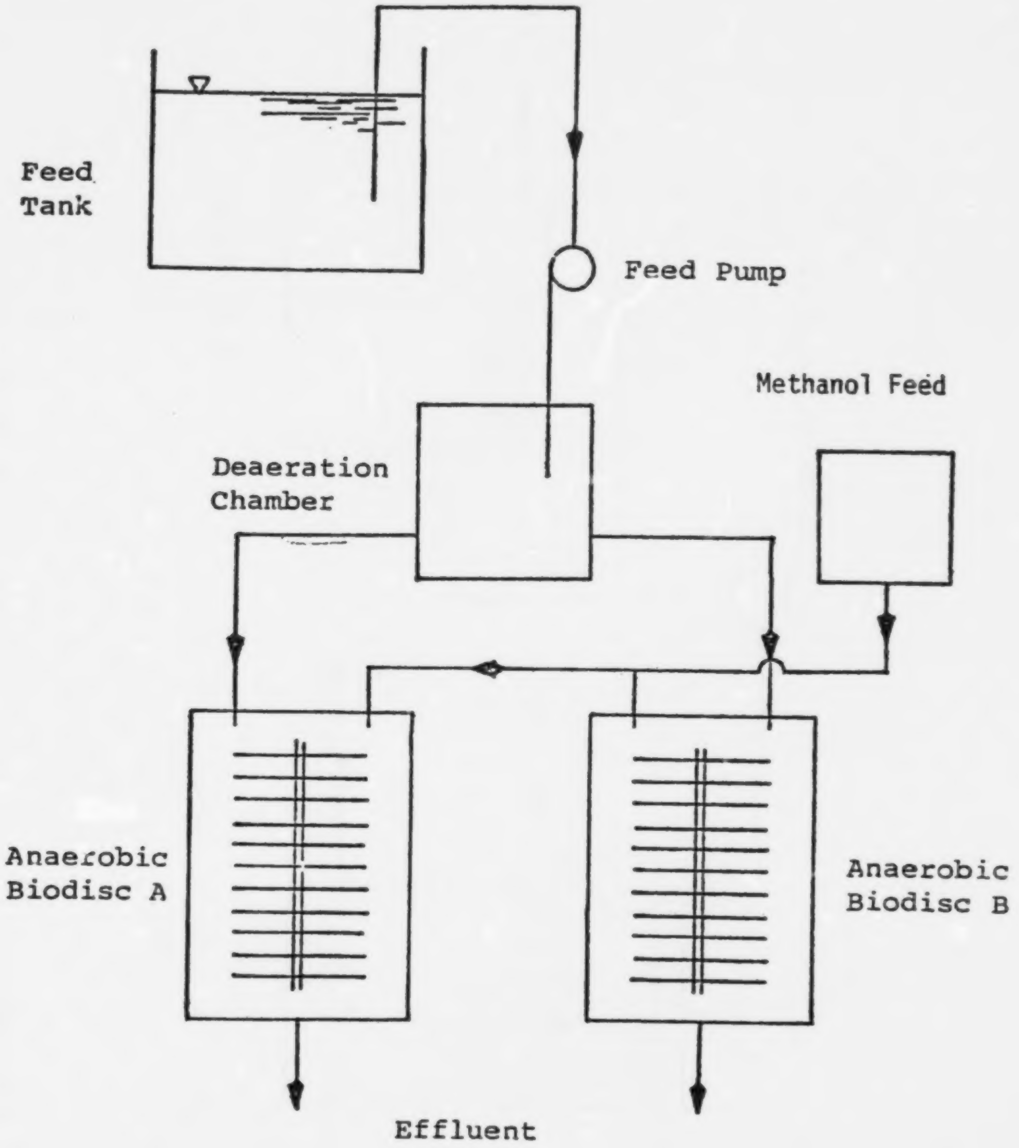

Feed Tank
Feed Pump
Methanol Feed
Deaeration Chamber
Anaerobic Biodisc A
Anaerobic Biodisc B
Effluent

FIGURE 2 : Flow Scheme for Laboratory Denitrification Units

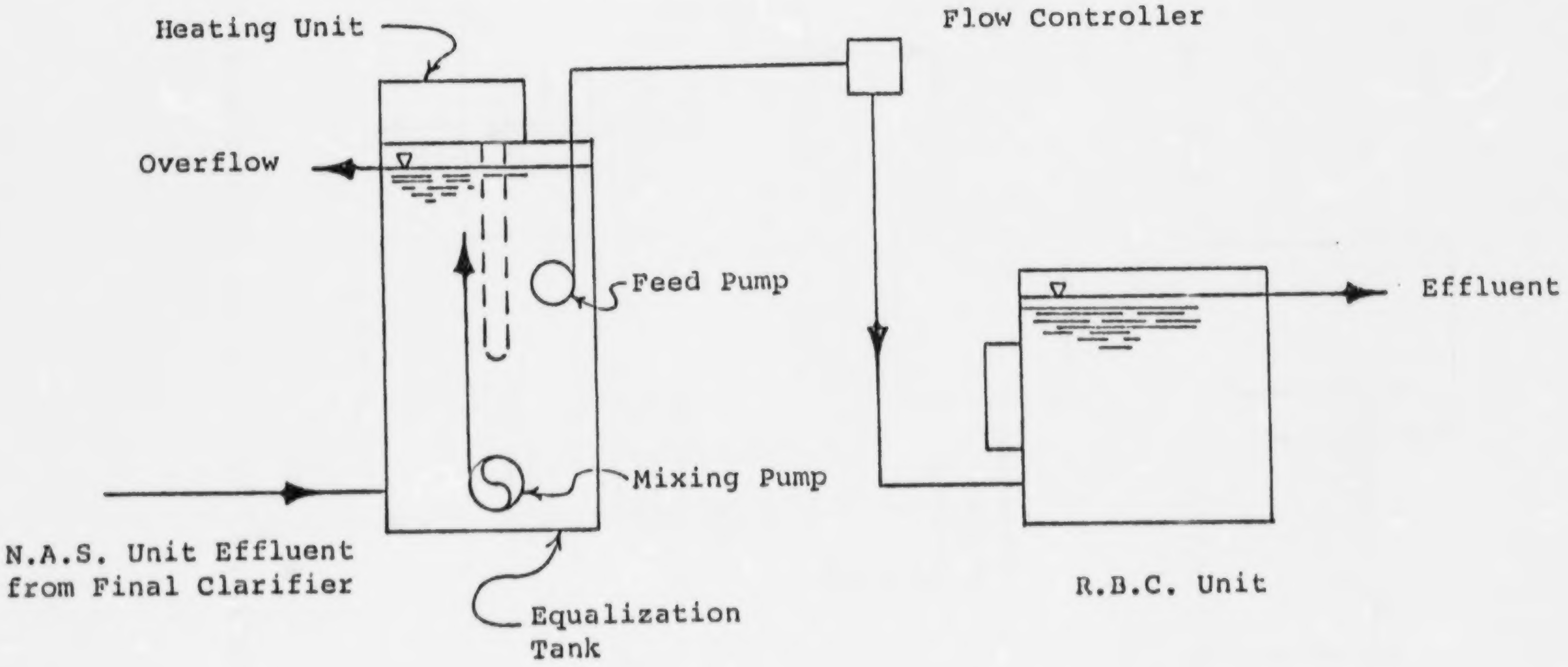

Heating Unit
Flow Controller
Overflow
Feed Pump
Effluent
Mixing Pump
N.A.S. Unit Effluent from Final Clarifier
Equalization Tank
R.B.C. Unit

Figure 3: Flow Scheme for Denitrification Pilot Study

Two airtight, single stage units were constructed of clear plexiglass. Each housed a stainless steel shaft supporting ten discs giving a total contact area of 9.08 ft/unit for biomass growth. Disc media was polyethylene, supplied by EPCO-HORMELL and designed to be 40% submerged. The wastewater volume of each unit was approximately .8 gallons, giving a fixed volume to surface ratio of 0.09 gal/ft^2. The units and a flow scheme are shown in Figure 2.

Influent was supplied via constant head peristaltic pumps through inert tygon tubing. Methanol, originally fed with an electrolysis pump, was later also supplied by peristaltic pumps.

Other unit components consisted of an influent deoxygenation chamber, in which N_2 gas was diffused through the wastewater, and an effluent clarifier designed to prevent air from entering the disc unit via the effluent line.

The single stage pilot plant portion of this project was conducted at the Marlborough Easterly Treatment Plant. A ten week study was conducted in the Fall of 1976 using clarified activated sludge from secondary tanks, which was further treated by pilot plant RBC nitrification units, as dentrification RBC unit influent. The unit was then washed down and reset for a 24 week study running directly from the final effluent (prior to chlorination) of the Marlborough Plant.

One single stage unit was utilized for the pilot study. A stainless steel shaft housed 12 polyethylene discs 2' in diameter, also of material supplied by EPCO-HORMELL, which combined with the wetted surface area of the unit provided 126.5 ft^2 for the attachment of microbial growth. The shaft was also attached to an external variable speed drive and rotated at 3 1/8 RPM under normal operating conditions. This unit was operated with the discs totally submerged in the wastewater which itself had a volume of 58.6 gal. The volume to surface ratio for this unit was 0.46 gal/ft^2. Figure 3 schematically illustrates the pilot unit operation.

A two stage denitrification pilot plant unit was operated at the Marlborough Easterly Treatment Plant from July 1977 to November 1977. A central stainless steel shaft held 20 polyethylene discs, 10 in each stage. The discs were 2 feet in diameter and were made of material supplied by EPCO-HORMELL. The spacing of the discs centerline to centerline was approximately one inch and each disc section was held in place by two plexiglass end pieces. The polyethylene discs provided 88.8 sq. ft. of surface area/stage for microbial growth. The surface area provided by the wetted walls of the unit and the plexiglass end pieces provided an additional 37.7 sq. ft. per stage for a total surface area of 126.5 sq. ft. per stage. The discs were submerged in a total volume of 69 gallons for each stage for a volume to surface area ratio of 0.546 gallons/ft^2.

The shaft was attached to an external variable speed drive and rotated at eight RPM under normal operating conditions. The unit was operated with the discs totally submerged in wastewater and the unit was covered.

ANALYTICAL METHODS

All samples were grab type samples. Analyses were either run immediately after collection, or the samples were refrigerated at 4°C. All analyses were completed within 24 hours of collection on the refrigerated samples.

TABLE 1: SUMMARY OF DENITRIFICATION IN BENCH SCALE UNITS

DATE	Nitrate Concentrations mg NO_3-N/l Inf	Unit A Eff	Unit B Eff	Percent NO_3 Removal Unit A	Unit B
5/10	23.0	2.0	---	91	--
12	25.0	1.8	---	93	--
18	20.0	1.0	---	95	--
20	20.0	1.4	3.6	93	82
6/02	20.0	2.3	8.0	88	60
6	11.0	1.1	5.0	86	45
15	13.2	0.2	0.4	98	97
17	23.0	0.4	---	98	--
21	20.0	0.0	2.8	100	86
29	16.0	---	0.6	--	96
7/01	13.6	0.5	0.5	97	97
3	12.4	0.4	0.2	97	97
7	14.4	0.2	0.8	99	94
8	15.2	0.3	0.3	98	98
9	15.6	0.6	0.6	96	96
13	21.6	0.4	0.3	98	99
14	18.4	0.3	0.4	98	98
Average Value	17.8	0.8	1.8	95	88

All laboratory analyses were conducted in accordance with the methods outlined in Standard Methods (12). Nitrate nitrogen was measured by the chromatropic acid method after Standard Methods (12).

EXPERIMENTAL RESULTS AND DISCUSSION

I - Laboratory Bench Scale Study

The denitrification bench scale study was run for two and one-half months, from early May through mid-July, 1976, in order to determine the feasibility of RBC denitrification and aid in pilot plant design. The hydraulic loading rate was held fairly constant throughout the study; however, the nitrate nitrogen concentration varied. The reason for the variation was that the feed influent used was the effluent from a nitrification RBC bench study.

The major problem encountered during the bench scale denitrification study was achieving the proper methanol addition to the influent, so a sufficient carbon source would be available to the denitrifying bacteria. In each case where either of the units did not achieve a high level of denitrification, there was insufficient methanol addition to the influent.

In evaluating the data, if insufficient carbon feeding was apparent, then that data were not used.

During the bench scale study, most of the loading rates were below 50 mg/m^2-hr (0.24 lbs NO_3-N/1000ft^2-day). Unit A had an average nitrate nitrogen effluent concentration of 0.81 mg/l with an average value of 95 percent of the nitrate nitrogen removed.

The influent averaged 1.9 mg TKN/l with the effluents averaging 1.3 mg TKN/l and 1.6 mg TKN/l for Units A and B, respectively. A summary of total denitrification performances is listed in Table 1.

The data indicated that a high degree of denitrification was achieved using rotating biological contactors. Over 90 percent of the nitrate nitrogen was converted to nitrogen gas, without any buildup in total nitrogen concentrations in the effluent. Near anaerobic conditions were maintained within the units and total nitrogen removal averaged greater than 85 percent.

II PILOT PLANT STUDIES

A - Single Stage Unit

A single stage denitrification pilot unit was run at Marlborough, Massachusetts Easterly WTP from October 1976 through June 1977. The unit was a single stage unit, which had the discs totally submerged in the wastewater stream, as described in a previous section of this paper. Methanol was used as the carbon source during this study.

During 1976 the denitrification unit was fed nitrified effluent from the RBC nitrification units. During 1977, the Unit was fed nitrified effluent for the Marlborough Easterly WTP activated sludge process. This effluent was collected prior to chlorination.

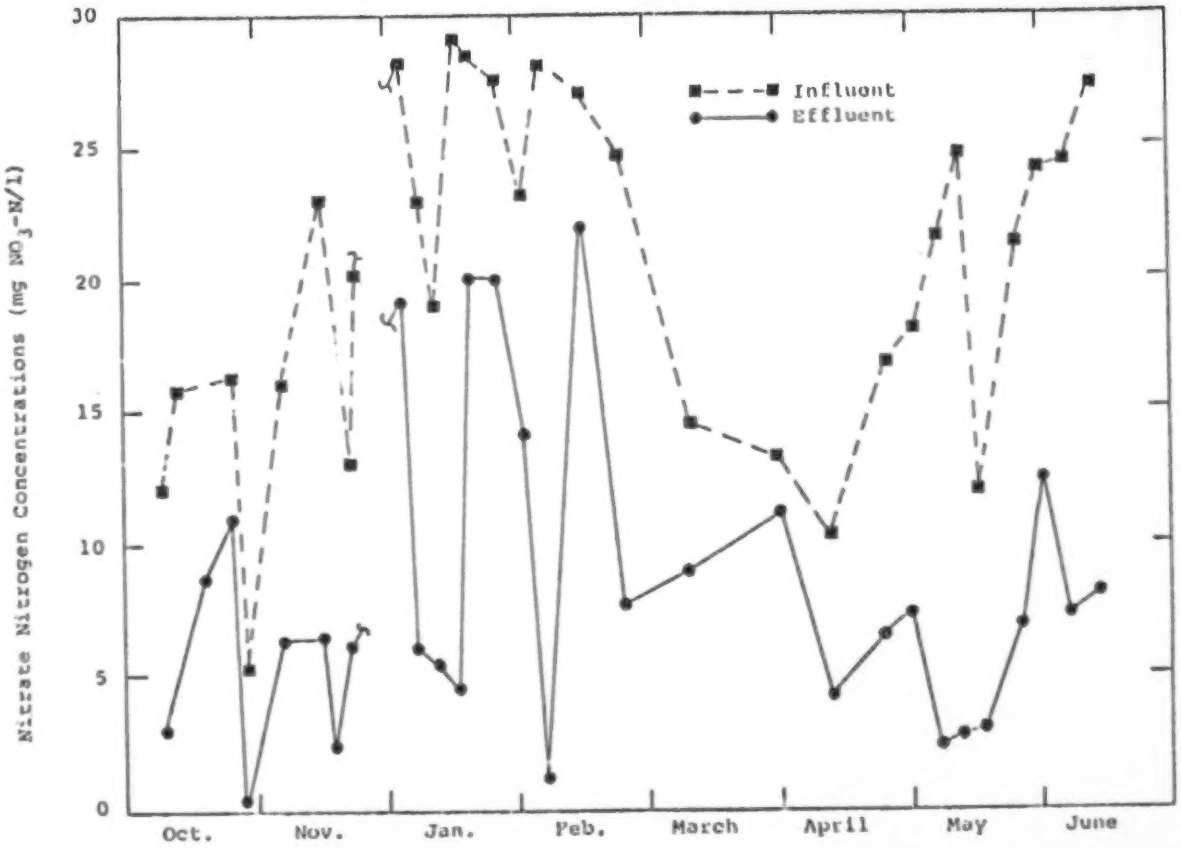


FIGURE 4; Nitrate Nitrogen Influent and Effluent Concentrations (Pilot Unit)

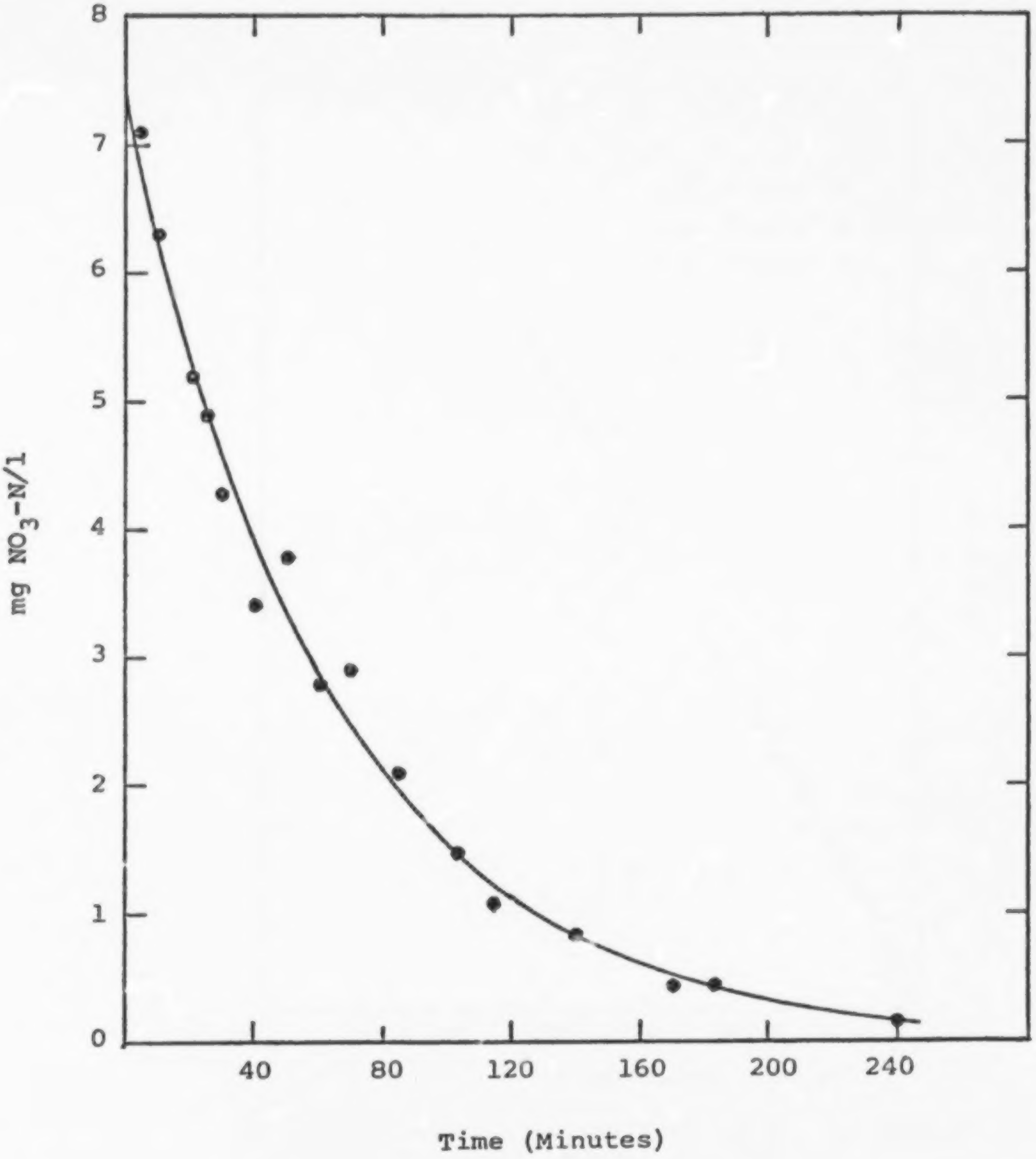


FIGURE 5: Nitrate Concentrations vs Time For No-Flow Conditions

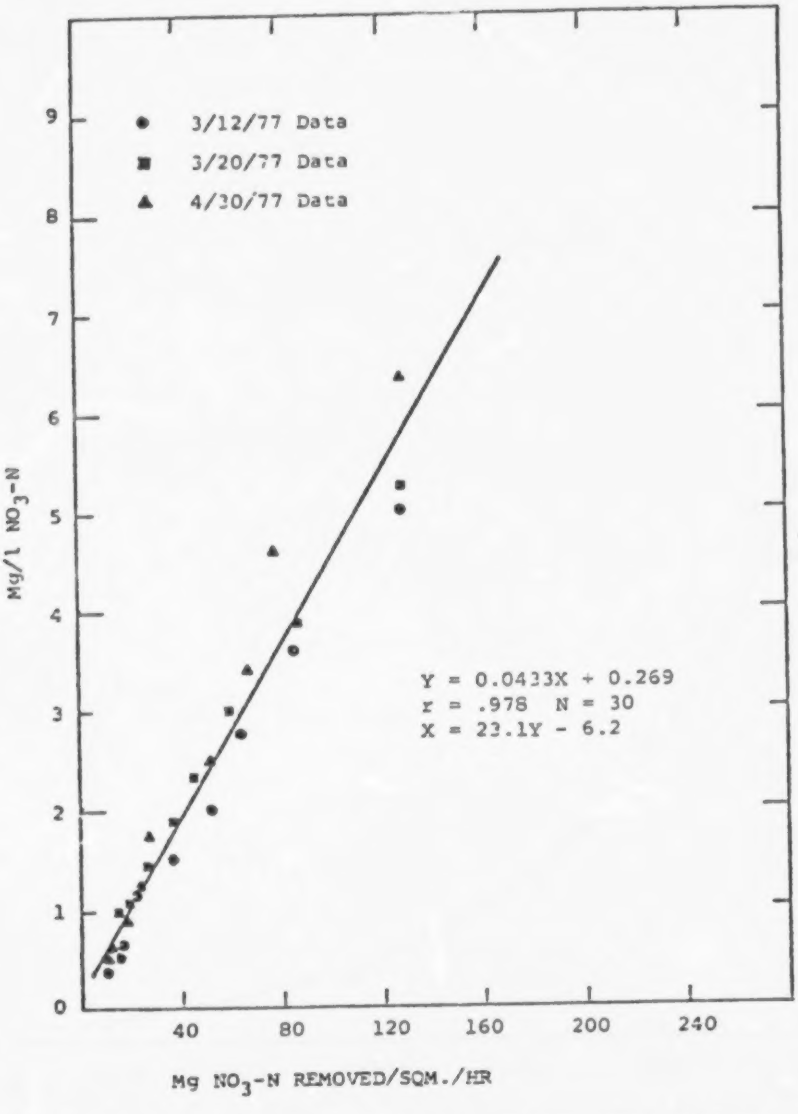


FIGURE 6: Nitrate Nitrogen Concentrations vs Nitrate Removal Rate for No Flow Conditions

Representative nitrate nitrogen concentration data collected throughout the study period are plotted in Figure 4. Because of three major factors:

1. the variability of the unit influent nitrate concentration;

2. the clogging problems in the pumping system, and thus high flow variations;

3. and the low nitrate nitrogen loading rates studied in the bench scale study,

loading rates on the pilot unit were varied throughout the study period. Generally, influent flow rates were held fairly constant for a minimum of one-week period. Steady state flow was achieved more often during the 1977 portion of the study.

Since the pilot unit was single stage, and the effects of staging could not be evaluated, higher loading rates and resulting removals were determined, rather than the lower loading rates studied with bench scale units.

During the operation of the pilot unit, it became apparent that the actual rate of nitrate removal was related to the nitrate nitrogen concentration within the unit. In order to evaluate the rate of reaction as affected by nitrate concentration, several batch tests were run. For these tests, the influent was shut off to the unit, and the nitrate level within the unit recorded with time. The results were parabolic curves of decreasing nitrate levels with time. A typical run is plotted in Figure 5. The slope of each curve at any nitrate level indicated the actual removal rate at that level. The data obtained from no-flow conditions was used to calculate the nitrate removal rates, and the rates were plotted vs. nitrate concentrations and shown in Figure 6. Fow low nitrate levels the reaction rates for the three dates tested are close to equivalent, however, at the higher nitrate concentrations, there was larger variation,

Based on this data, the removal rate, at any oeprating level of nitrate, within the unit, was calculated to be:

$$R = 23.1\ N - 6.2$$

where: R is the mg/sq m/hr removed and

N is the nitrate level mg NO_3-N/1 in the unit.

Being a complete mix system, the effluent level would also be the same as that within the unit.

The correlation coefficient for this curve is 0.98 and therefore, over 95% of the variation in the rate of removal is accounted for by the change in the nitrate level within the reactor unit.

As the RBC is designed to be a continuous flow unit, a similar comparison with flow-through conditions was necessary.

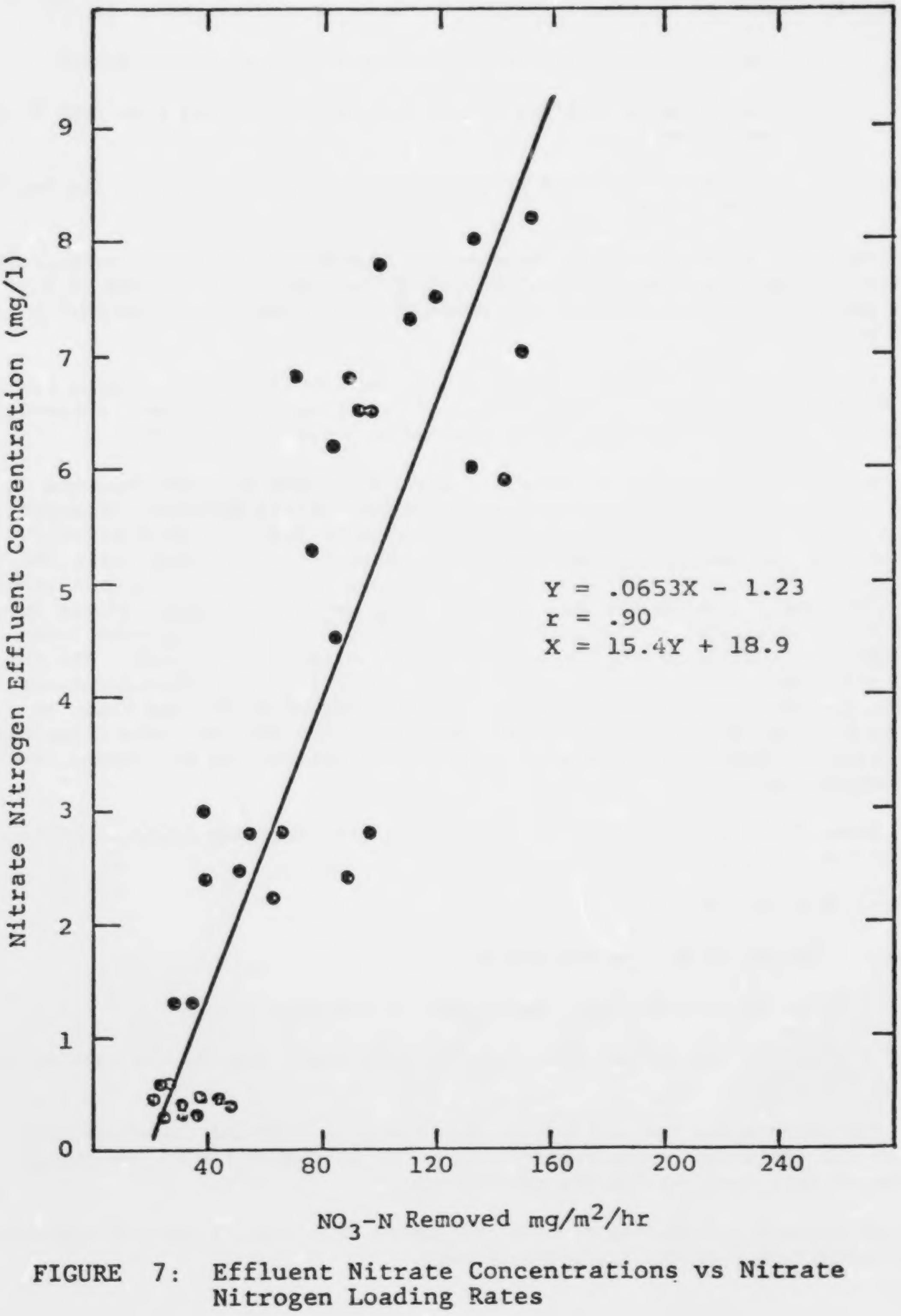


FIGURE 7: Effluent Nitrate Concentrations vs Nitrate Nitrogen Loading Rates

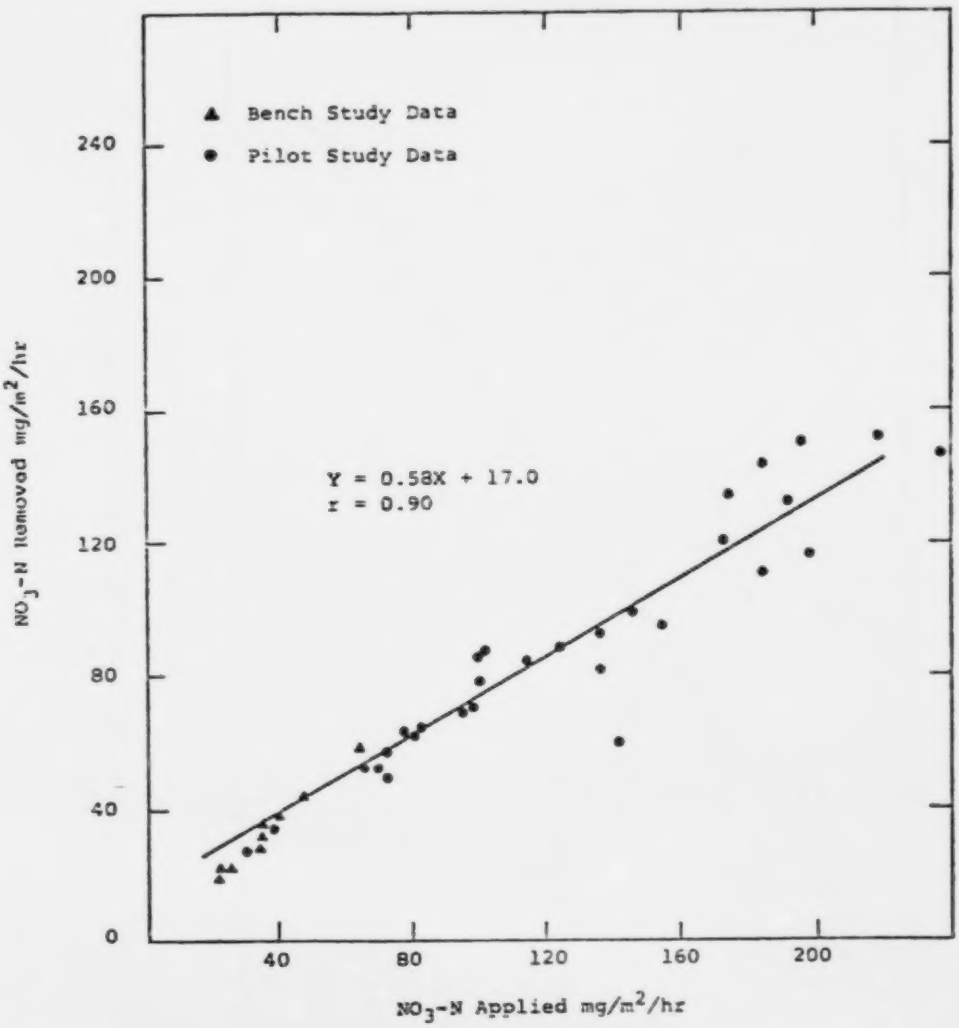


FIGURE 8: Nitrate Nitrogen Removed vs Nitrate Application Rates

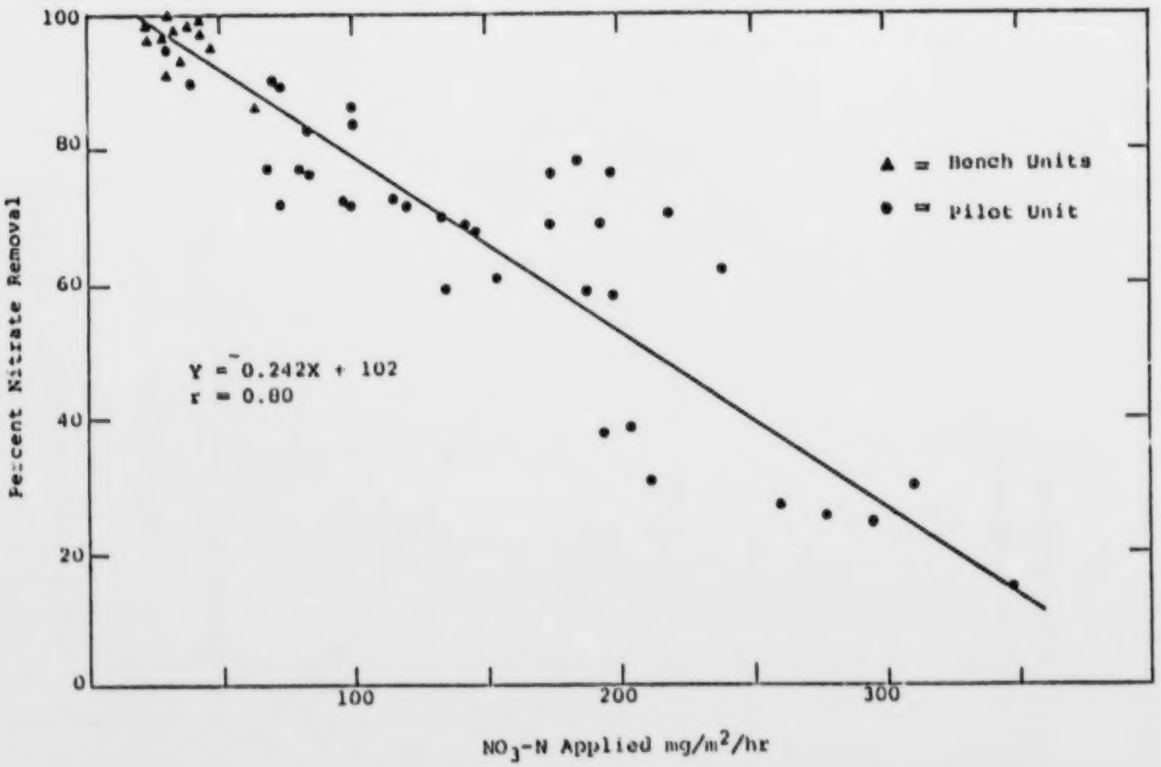


FIGURE 9: Percent Nitrate Nitrogen Removed vs Nitrate Application Rates

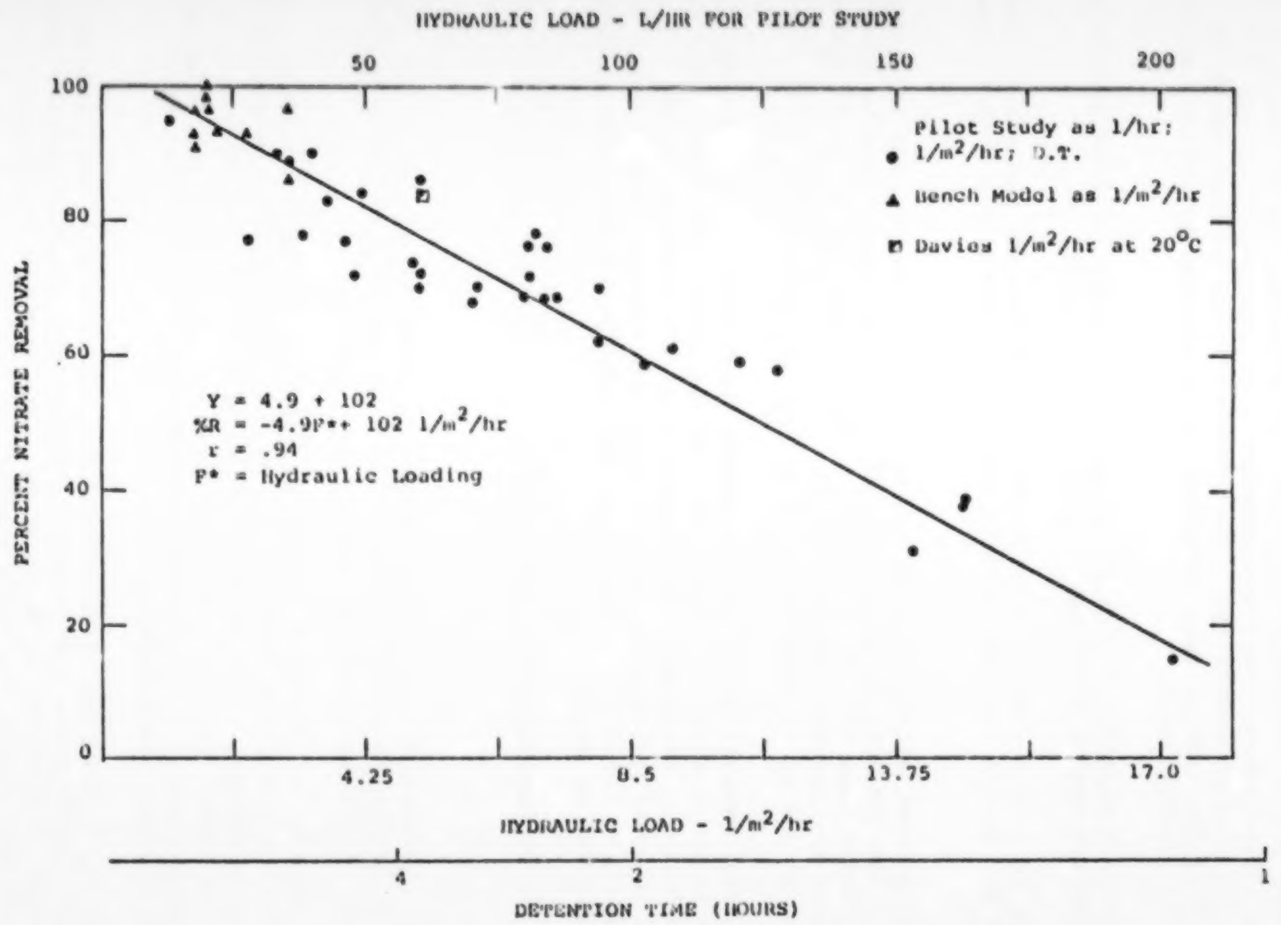


FIGURE 10: Percent Nitrate Removed vs Hydraulic Loading Rate

With the unit operating at various flow rates, temperatures, and nitrate influent and effluent levels, actual removal rates were determined. Although some oxygen entered the unit, and at times was not completely removed, its effects were not initially considered. Figure 7 shows the nitrate nitrogen removal rate vs, the nitrate nitrogen concentrations in the effluent for data collected in a flow-through situation. As expected, due to the varying conditions listed above, there was somewhat more scatter within the data than when under no-flow conditions. The correlation coefficient was 0.90, such that 81% of the removal rate variation could actually be attributed to the change in nitrate level.

The data plotted in Figure 7 neglect correction for both temperature and any dissolved oxygen which may be present in the unit.

Figure 8 is a plot of nitrate nitrogen removal for various nitrate loading rates. As the influent nitrate loading increased, the nitrate removal decreased.

The percent nitrate removal vs. nitrate application rate is plotted in Figure 9 and the data indicate that 90% nitrata-N removal can be expected at loading rates less than 55 mg/sq m/hr. The removal may be approximated from the linear relationship shown on Figure 8. This relationship was for loading rates up to 250 mg NO_3-N/m^2-hr. or 1.22 lbs. NO_3-N/1000ft^2-day.

Nitrate removal data from the single stage unit was plotted vs. hydraulic loading rates (l/m^2-hr) in Figure 10. As the hydraulic loading, and thus the nitrate loading, increased, the percent nitrate removal decreased. Hydraulic loading rates of 8.5 l/m^2-hr. had at least 60% nitrate removal while hydraulic loadings below 2.0 l/m^2-hr. had 90 to 95% nitrate removal. The relationship was calculated as being:

% Nitrate Removal = 102 - 4.9 Q,

where Q = l/m^2/hr.

The plotted data was not corrected for temperature, and was in good aggreement with Davies' data (10). Again this data was obtained using a single stage unit with variable nitrate and temperature values.

In this study, methanol (CH_3OH) was used as the carbon source. Varying amounts of methanol were added to the wasteflow in order to determine optimum carbon requirements. This data can be found in detail in reference (13). The results indicated that in a single stage unit, addition of more than 80 mg COD/l, carbon addition did not increase nitrate removal efficiencies, irrespective of the nitrate loading rate,

Figure 11 shows the ratio of methanol addition to nitrate removal for different loading rates and removal efficiencies. The nitrate application rates ranged from 40-180 mg NO_3-N/m^2-hr, The methanol requirement is based on the COD requirement for NO_3-N removal and D.O, removal within the unit. For this study, the methanol requirement ratio in terms of COD was calculated to be 1.37. The ratios shown in Figure 11 are for CH_3OH requirements (not COD) added to achieve NO_3-N removal. These ratios are for influent nitrate

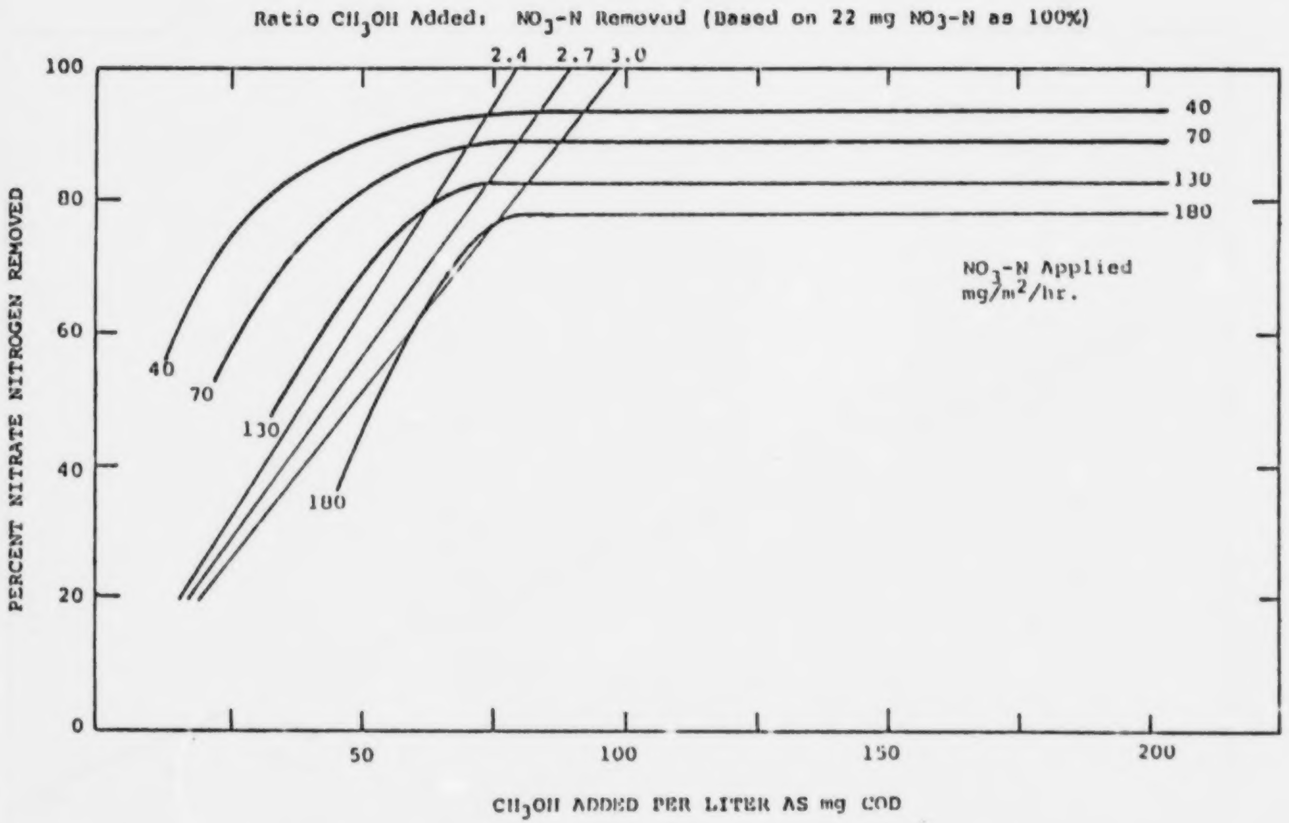


FIGURE 11: Effects of Methanol to Nitrate Ratios on Percent Nitrate Removals

FIGURE 13: First Stage Removal Efficiency

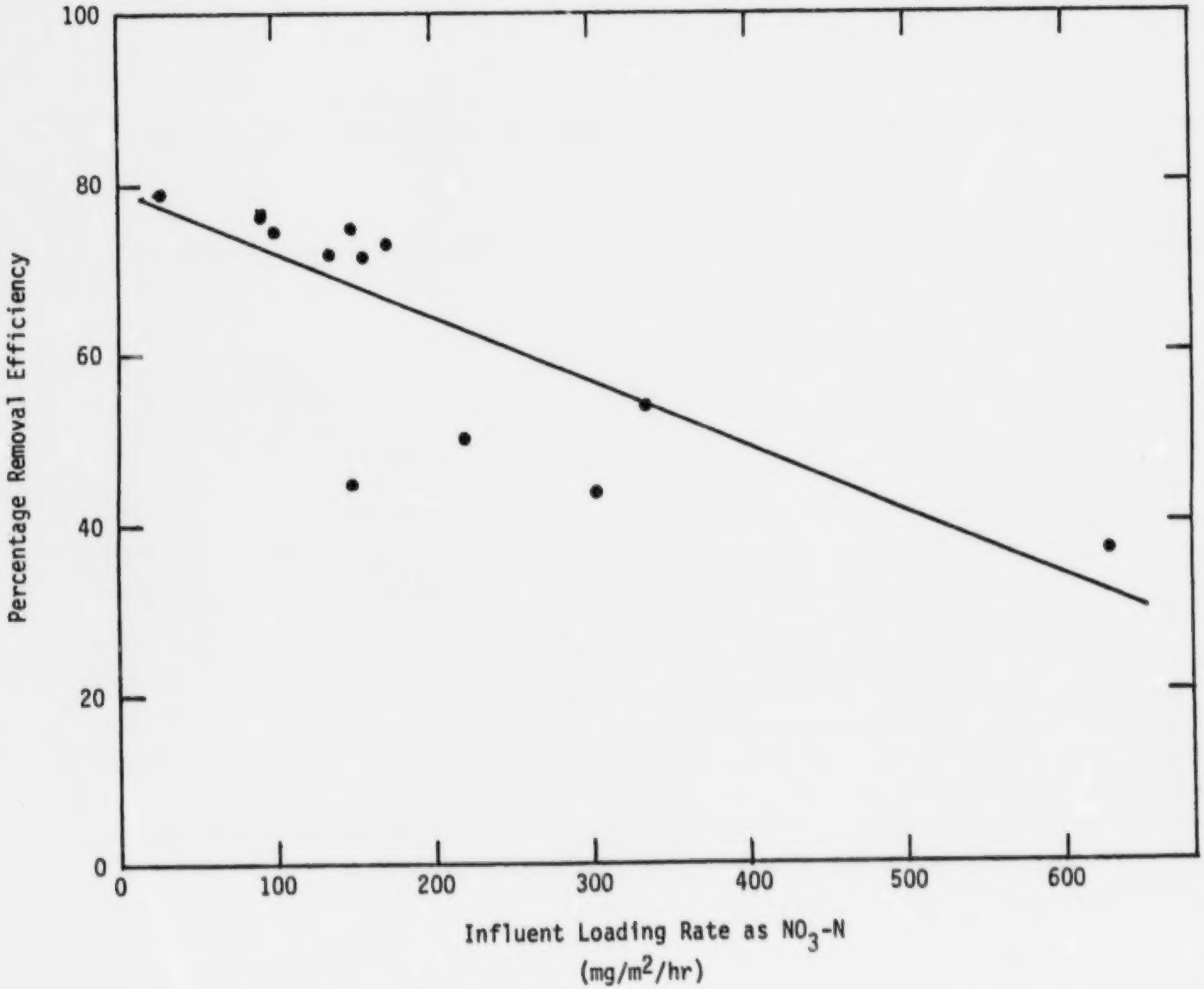


FIGURE 12: First Stage Nitrate Removal

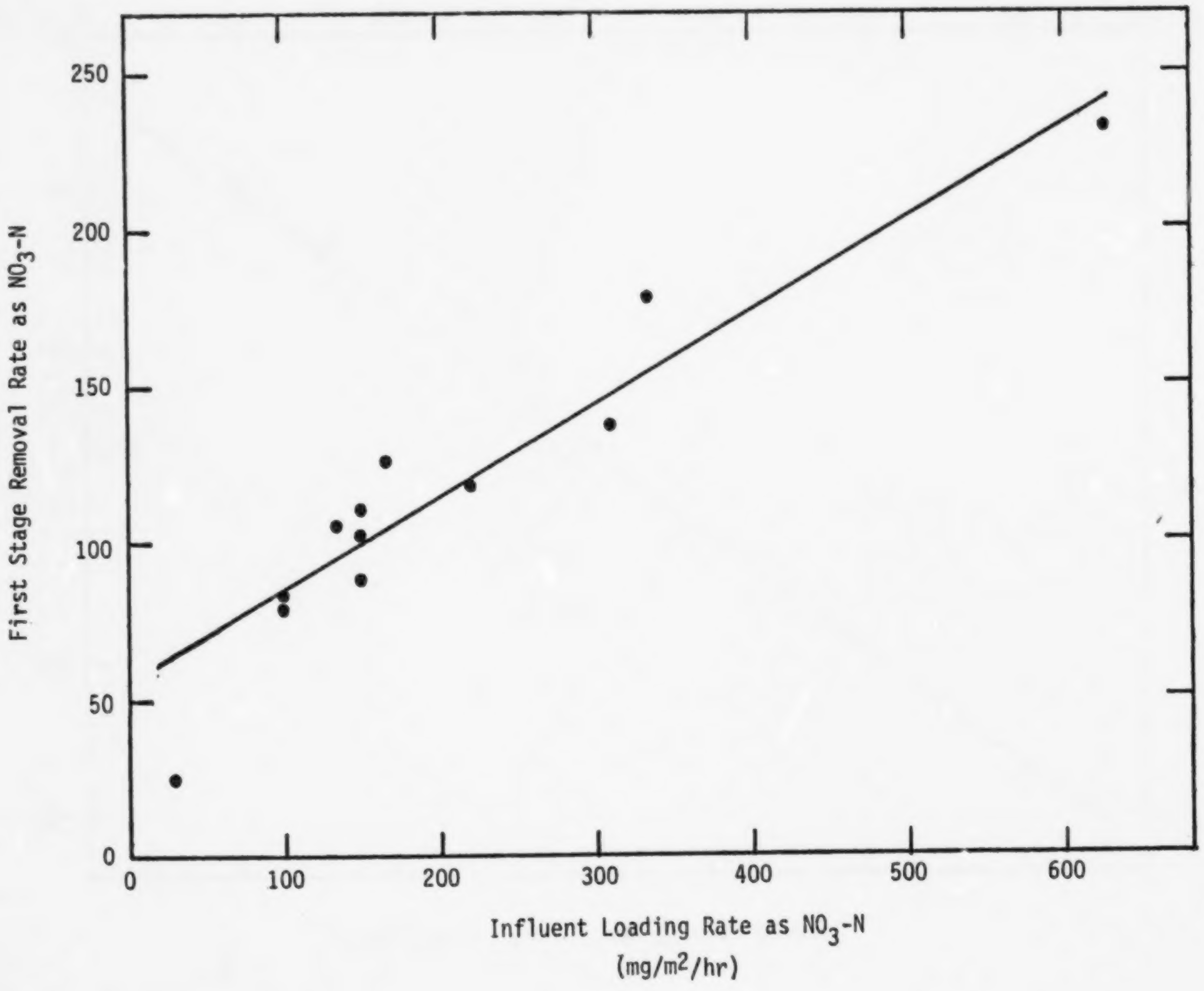


FIGURE 14: Two Stage Nitrate Removal

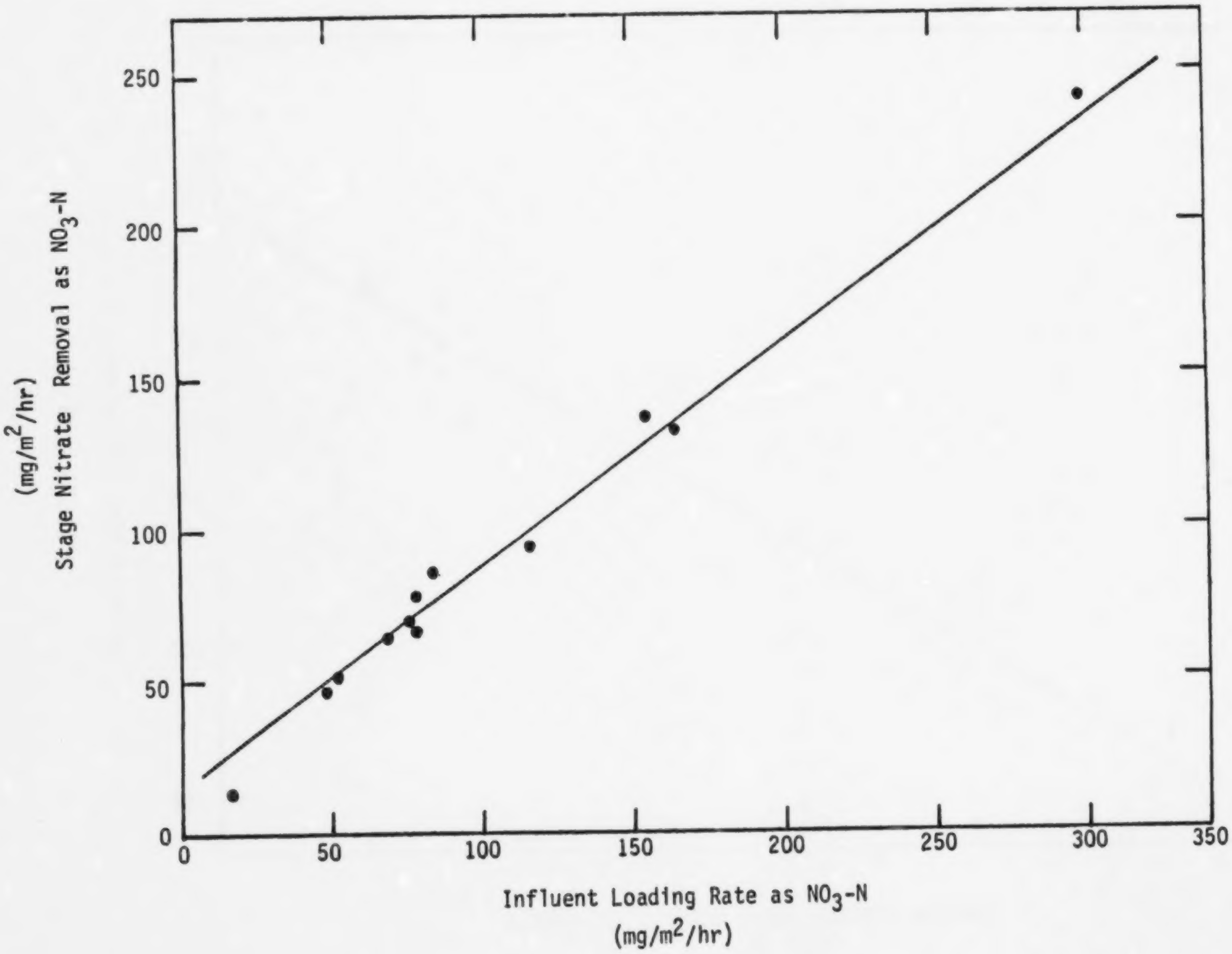
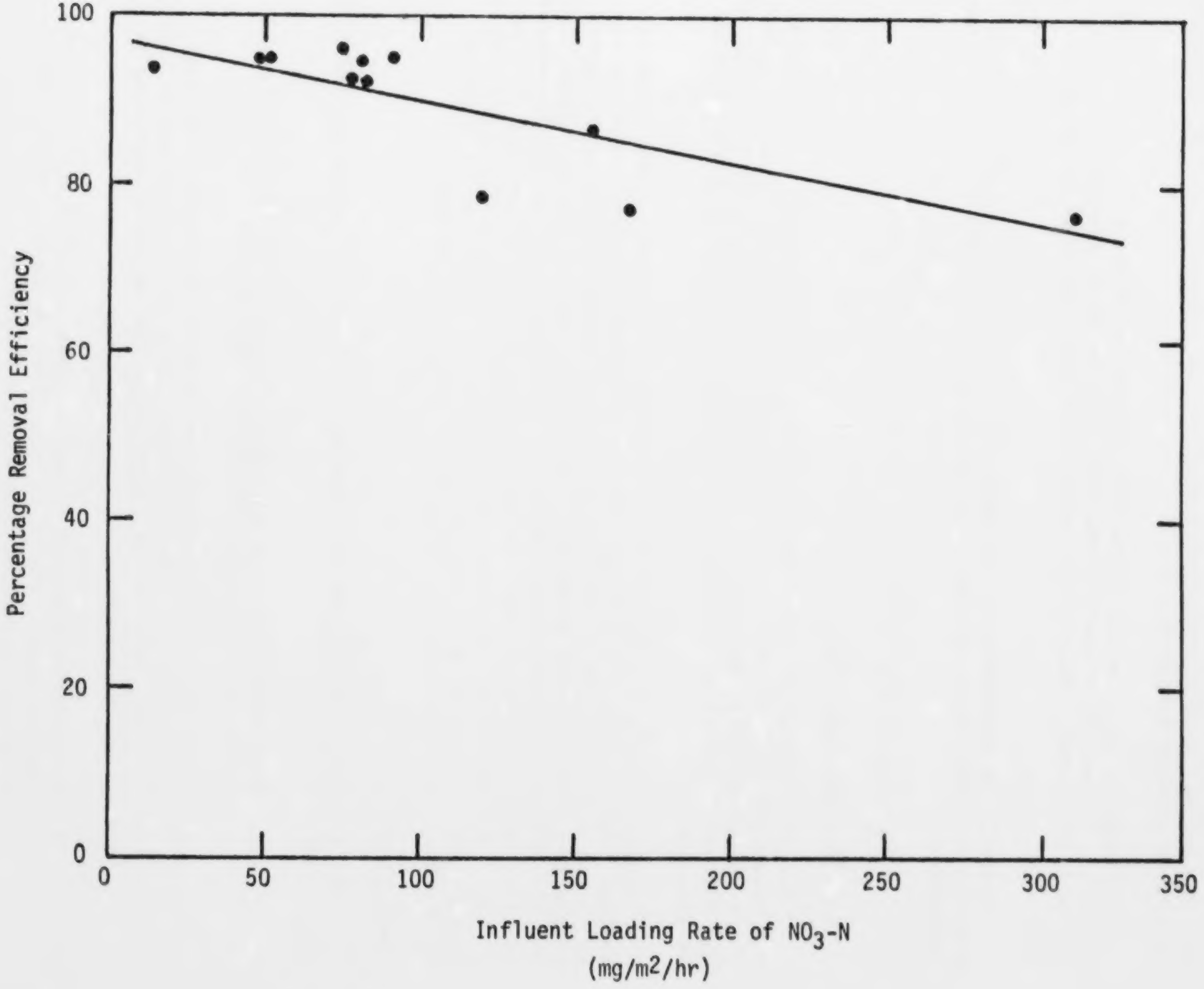


FIGURE 15: Two Stage Removal Efficiency



nitrogen concentrations of 22 mg NO_3-N/1.

The data shown in Figure 11 indicates that the nitrate removal efficiencies in the RBC unit was a function of both nitrate loading rate, and the carbon to nitrate ratio used. In this study, for the loading rates investigated, the optimal methanol to nitrate ratio varied between 2.4 - 3.0, depending on nitrate loading rate.

Although an exact solids production monitoring balance was not continually conducted, the solids production was low. This is to be generally expected in anaerobic processes. The average value of effluent suspended solids was 3.4 mg/1.

Temperature dependency in the range of 6 to 16°C was also evaluated in this study. The results indicate that it could be described by the following expression:

$$K_T = K_{20} \left[1.075^{(T-20)}\right]$$

B - Two Stage Pilot Unit

The two stage pilot plant was operated at the Marlborough W.T.P. during the summer and fall of 1977, one year after the single stage operation. Each stage was essentially the same in surface area and volume as the one stage unit. his unit was operated at a higher rotational speed and had a greater spacing between discs.

Figures 12, 13, 14, and 15 show the curves of nitrate loading rates vs. removal rates and nitrate loading rates vs. percentage removal rates for the two stage unit. Figures 12 and 13 present first stage results while Figures 14 and 15 present the two stage unit results. To compare the effects of staging for the two stage investigation in 1977, Figures 13 and 15 may be examined simultaneously. These figures indicate that at an application rate of 1 lb. of NO_3-N per 1000 ft^2 of surface area, (205 mg NO_3-N/m^2-hr), the single stage would remove 63 percent of the nitrate while a two stage unit would remove 83 percent. In both instances, the contact times were the same.

CONCLUSIONS

The following conclusions were reached based on the studies described in this paper:

1. Denitrification of a nitrified wastewater can be readily accomplished using rotating biological contactors.

2. There is a linear relationship between the amount of nitrate applied and the amount removed. This can be expressed by the equation; NR = .59 NA + 18.8, with the variables expressed as mg NO_3-N/sq m-hr.

3. Nitrate concentration in the unit has a significant effect on removal rates. The higher the level of nitrate within the unit, the higher the removal rate will be.

4. 90% nitrate reduction can be expected at loading rates of up to 55 mg NO_3-N/sq m-hr under normal operating conditions in a single stage reactor. For a two stage reactor, 90% nitrate reduction will occur at loading rates up to 100 mg NO_3-N/sq m-hr.

5. Removals will increase with increasing detention times for any unit at a given influent nitrate concentration.

6. Methanol requirements for optimum removal can be estimated as: CH_3OH mg/l = 2.6 NO_3-N mg/l + 0.91 D.O. mg/l. Lesser amounts limit removals and excess amounts produce high effluent BOD values.

7. The temperature dependency can be estimated as:

$$K_T = K_{20}\ 1.075^{(T-20)}$$

8. Solids production was low at all loading rates during these investigations.

9. Staging will improve the performance of RBC denitrification units.

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NITROGEN REMOVAL IN ROTATING BIOLOGICAL CONTACTORS WITHOUT THE USE OF EXTERNAL CARBON SOURCE

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Introduction

The main pollution problem resulting from wastewater discharges in Norway is eutrophication of the inland lakes and the fjords. In order to fight eutrophication, phosphate removal is now carried out in most treatment plants. The role of nitrogen in eutrophication has been more or less neglected, and treatment plants for nitrogen removal are non-existant in Norway. The interest in nitrogen removal is however growing caused by the fact that nitrogen has been found to be growth limiting nutrient in some eutrophic lakes.

The interest for nitrogen removal combined with the rising popularity of the biodisc in Norway, has initiated research at The University of Trondheim on the use of biodisc-plants for the removal of both organic matter, phosphorus and nitrogen.

NITROGEN REMOVAL IN BIODISC-SYSTEMS

The literature on biological nitrogen removal in general is vast, at least compared to the number of actual plants that have been built throughout the world. The majority of the papers on nitrification/denitrification, however, deals with suspended cultures. Most of the denitrification studies that have been carried out with fixed film cultures, have dealt with anaerobic filters or lately, fluidized beds. In the last few years some authors have reported from investigations of denitrification in biodisc systems /1, 2, 3, 4/.

The flow diagram normally used when biodiscs are to be used for both nitrification and denitrification, is the one shown in Fig. 1.

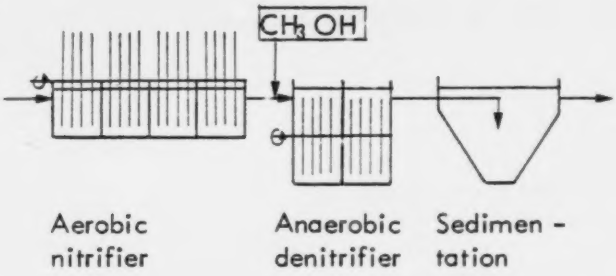


Fig. 1. Normal process for nitrogen removal in biodisc plants.

The advantage of this system compared to a suspended culture system or to a system with anaerobic filter as denitrifier, is that only one sedimentation unit is needed.

The main disadvantage of the system, is the need for an additional carbon source to obtain denitrification. As in other denitrification systems, methanol would normally be used. There are several disadvantages with methanol addition:

1. It increases operation costs. The enormous increase in methanol price the last few years, has made this point very important.
2. It increases investment cost since an advanced control system is necessary in order to add the correct amount of methanol.
3. Surplus methanol not used for denitrification would increase the effluent BOD. The amounts normally added corresponds to 80 - 100 mg BOD/l in municipal sewage, which could be detrimental to the receiving water if denitrification for some reason did not take place or in case of failure in the control system mentioned above.

The goal with our research was to unnecessitate the need for an external carbon source and use the raw water itself as such. This has successfully been done in suspended culture systems in the so-called Bardenpho-process proposed by Barnard /5/, where nitrate-rich mixed liquor from an activated sludge tank, is recycled to an anaerobic suspended culture tank ahead of the aerobic one.

In the process presented in this paper, the same principle has been utilized in a biodisc-system, see Fig. 2.

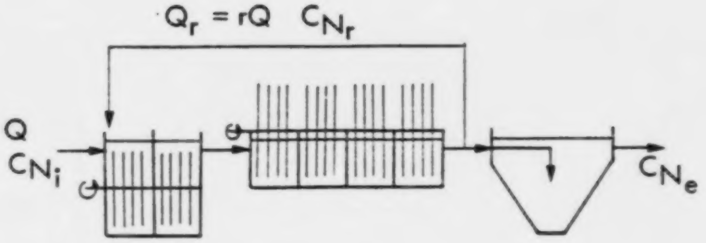


Fig. 2. Proposed process for nitrogen removal in biodisc plants.

The ammonium-rich influent passes through a tank with submerged biodiscs where anoxic conditions will exist. From this tank it enters an ordinary aerobic biodisc unit where nitrification takes place. Nitrified water is then recycled to the inlet of the anoxic tank where denitrification takes place and where raw water can act as carbon source.

The maximum possible nitrogen removal that can be achieved with the process, may be calculated from a mass balance assuming complete nitrification and denitrification in the two steps:

$$Q\,C_{N_i} = rQC_{N_r} + QC_{N_e}$$

$$= rQC_{N_e} + QC_{N_e} = QC_{N_e}(1+r)$$

Removal:

$$\frac{C_{N_i} - C_{N_e}}{C_{N_i}} \cdot 100\% = \frac{C_{N_i} - \frac{C_{N_i}}{r+1}}{C_{N_i}} \cdot 100\% = \frac{r}{r+1} \cdot 100\%$$

where C_{N_i}, C_{N_e} and C_{N_r} are nitrogen concentrations in the influent, in the effluent and in the recycled flow respectively.

A removal of 90% requires for instance a recycled flow of 9 times the influent flow, while 80% is achievable with a recycle ratio of $r = 4$.

EXPERIMENTAL ARRANGEMENT

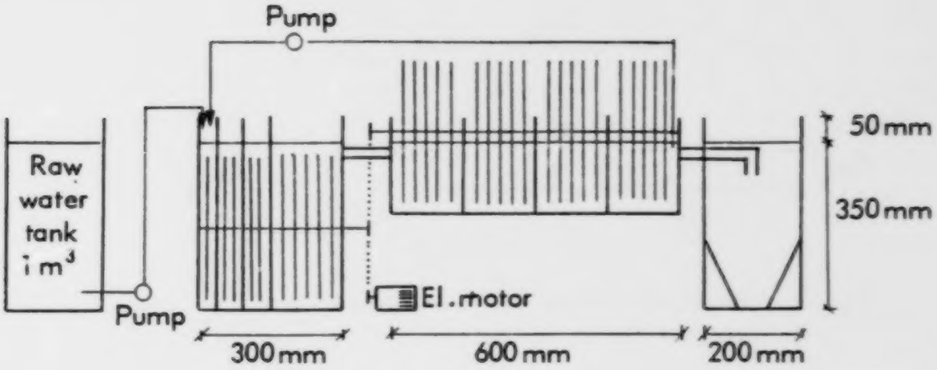
The experiments were carried out in two plexi-glass biodisc units arranged in parallel, each with a capacity of 1 - 10 l/h (Fig. 3).

Experiments with three different types of raw water have been carried out:

1. Artificial wastewater
2. Municipal wastewater (sewage)
3. Leachate from sanitary landfill

Artificial wastewater

The reason for starting out with artificial wastewater was to ensure influent characteristics to be as stable as possible in order to investigate the possibilities of the process. The water was composed to be as similar to municipal wastewater as possible.



	Denitrifier	Nitrifier
Disc diam.	300 mm	300 mm
Disc number	10	20
Comp. number	4	4
Disc area	1.41 m^2	2.82 m^2
Tank volume	32.8 ℓ	23.6 ℓ
Rotational speed	15-16 rev/min	15-16 rev/min

Fig. 3. Experimental unit.

Table 1 Composition of artificial waste water

Glucose	200 mg/l	$KHCO_3$	50 mg/l
Yeast extract	30 "	$NaHCO_3$	130 "
Dried milk	120 "	$MgSO_4 \cdot 7\,H_2O$	50 "
Urea	30 "	$FeSO_4 \cdot 7\,H_2O$	2 "
NH_4Cl	60 "	$MnSO_4 \cdot H_2O$	2 "
$Na_2PO_4 \cdot 12\,H_2O$	100 "	$CaCl_2 \cdot 6\,H_2O$	3 "
		Bentonite	40 "

This water gives the following results when analyzed:

Total COD (COD_t) : 400 mg O/l
Soluble COD (COD_s) : 300 "
Soluble BOD (BOD_{5_s}) : 188 "
Tot P : 9 mg P/l
Tot N : 36 mg N/l
Suspended solids : 90 mg/l
Alkalinity : 130 mg $CaCO_3$/l

Sewage

The sewage was taken from an apartment house during morning hours in order to have high ammonium concentrations. The water showed, however, after presettling, to have a relatively low soluble organic content ($COD_{filtered}$ ≈ 100-160 mg O/l) compared to the ammonium content (NH_4-N ≈ 35 - 65 mg N/l). As will be commented later, this gave us some problems since COD became the limiting factor for denitrification. In part of the experiments this relationship between COD and N was corrected for by introducing dairy effluent (dried milk) to the sewage.

Leachate from Landfill

The high amount of ammonium in leachate may cause both oxygen deficit and eutrophication in the receiving waters, and a couple of such situations in Norway have intensified research on treatment of leachate from landfills.

A review through the literature /6/ will show that treatment of leachate is very troublesome, and that anaerobic treatment has been used with some success. We found it interesting, therefore, to test the proposed process on leachate.

The leachate used in this experiment was that of the landfill of Trondheim. The experiments were carried out during a period of heavy rainfall that diluted the leachate. Table 2 shows composition of the leachate during 1975 and 1976 compared with the composition during our experiment.

Table 2 Composition of leachate

Parameter	1975 - 76	Actual experiment
COD mg/l	1250 - 1650	400 - 800
Tot N	80 - 100	80 - 130
NH_4-N	65 - 80	70 - 125
Tot P	0.7 - 0.8	0.6
Fe	105 - 140	100 - 145
pH	6.8	7.0

The difference is especially significant for COD while the difference for the other parameters are minor.

In order to prevent the problems with ironhydroxide-precipitation on the biofilm, most of the iron was precipitated in advance by adjusting the pH to 8, aeration and filtration.

RESULTS

Nitrogen removal

In Fig. 4 are shown experimental results from situations where nitrification and denitrification limited by any factor. The results are compared with the theoretically possible removal as mentioned earlier.

The figure shows that the results fit the theoretical removal quite nicely. This demonstrates that given the situation that no factor (such as pH, O_2-concentration, BOD-loading, toxic substances, and so on) limits nitrification or denitrification, the proposed process will work as anticipated. Concequently, the process is an interesting alternative to the different nitrogen removal processes used today.

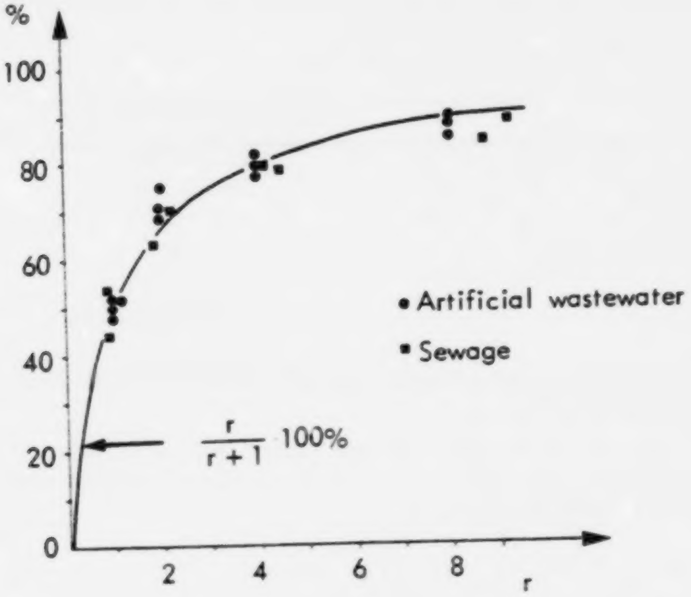


Fig. 4. Nitrogen removal against resirculation ratio.

Nitrification

Several factors are influencing the degree of nitrification, such as content of biodegradable organic matter, oxygen, pH and temperature.

Most workers agree that the oxygen content in the water may be a limiting factor with regard to the nitrification rate. Weng and Molof /7/ found that nitrification took place only if more than 2 mg O_2/l was present. In our experiments the oxygen content was normally in the range of 3-6 mg O_2/l and never below 2.5 mg O_2/l in the nitrifier.

Optimum pH for nitrification has been found to be around 8 /8/, but nitrification is achieved in the range of pH = 6.5 - 9.5. In all our experiments the pH was in the range of 7.0 - 8.3.

Optimum temperature for nitrification has been found to be in the range of 28 - 36 °C /9/ where municipal treatment processes seldom operate. Antonie /2/ has postulated that the nitrification rate is constant over 15 °C, but decreases with decreasing temperature under that. Pretorius /10/ found Q_{10} $(= \hat{\mu}_{20}/\hat{\mu}_{10}) = 1.25$ in his biodisc experiments.

No temperature control was used in the experiments presented here. During the artificial waste water period, the temperature was in the range of 13 - 16 °C, while it varied more in the sewage and the leachate period (10 - 16 °C).

In order to achieve nitrification, the content of biodegradable matter has to be brought down sufficiently.

Prior to the experiments with the proposed treatment process (Fig.2), some experiments with the process shown in Fig. 1 were performed in order to investigate the acceptable organic loading to achieve nitrification. These experiments were carried out with the artificial wastewater.

Fig. 5 shows the relationship between the degree of nitrification (per cent removal of NH_4-N) and the content of soluble organic matter ($COD_{filtered}$) in the effluent.

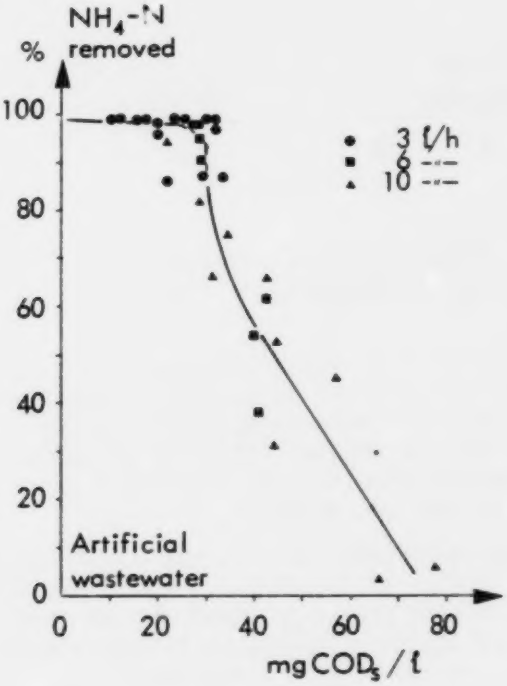


Fig. 5. Effect of soluble COD on the degree of nitrification.

It is demonstrated that a soluble COD-content of less than 30 mg COD/l was needed to ensure complete nitrification. Antonie /2/ reported that nitrification started when the BOD-concentration was brought down to 30 mg BOD_5/l, while it was fully developed below 8 - 10 mg BOD_5/l. Weng and Molof /7/ found that nitrification occurred at COD-concentrations below 50 mg COD/l. Both of these findings correspond to a soluble COD-concentration of about 25 - 35 mg COD/l and are in agreement with our results.

In Fig. 6 is plotted the degree of nitrification against organic loading of soluble COD per unit area of biofilm. It is demonstrated that the load had to be kept below 25 g $COD_S/m^2 \cdot d$ to ensure full nitrification.

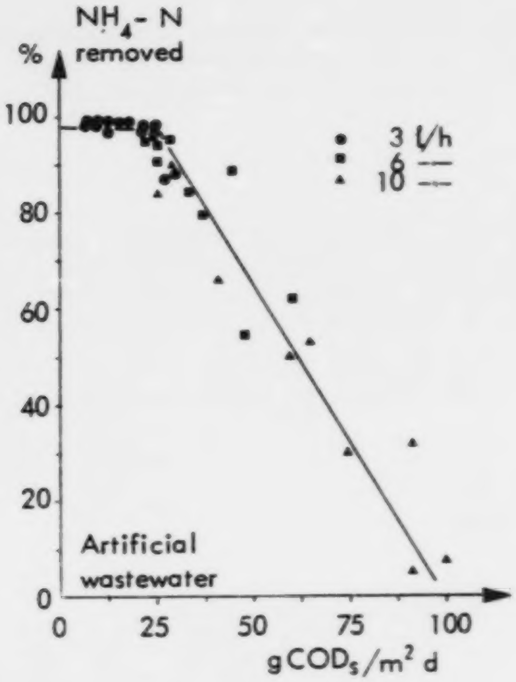


Fig. 6. Effect of organic loading on the degree of nitrification.

To evaluate the results on nitrification in the experiments with the proposed anoxic-aerobic process is a little complicated because the recirculated water contributes both to the COD-loading and the ammonium-loading. The COD that enters the nitrifier is probably composed of one easy biodegradable fraction from the raw water and one heavy biodegradable fraction from the recirculated water.

In the sewage experiment the BOD_5 in the recirculated water for one run was found to be as low as 3.5 mg BOD_5/l, while the COD was 39 mg COD/l. It may thus be stated that the organic load from the recirculated water does not contribute much to the total biodegradable organic load that will determine the degree of nitrification.

In the experiments presented in Figs. 5 and 6, recirculation was not used. In the experiments with the proposed process higher COD_s-values could be tolerated to ensure nitrification. This is partly due to the fact that non-biodegradable COD is recirculated. In the leachate experiments, nitrification took place even at COD-values of 150-200 mg COD/l.

In Fig. 7 is plotted the nitrification rate against ammonium-loading for situations where nitrification took place (was not limited by organic loading). The spread in the results may be explained by the lack of temperature control.

It is interesting to note that for all the types of raw water tested, there seems to be a critical loading of about 100 - 150 mg NH_4-N/m^2·h over which the nitrification rate is at its maximum and independent upon the loading. This maximum nitrification rate is about 100 - 120 mg NH_4-$N_{removed}$/m^2·h both for the artificial wastewater and the sewage, while it is considerably lower for the leachate, around 40 mg NH_4-$N_{removed}$ per $m^2 \cdot h$.

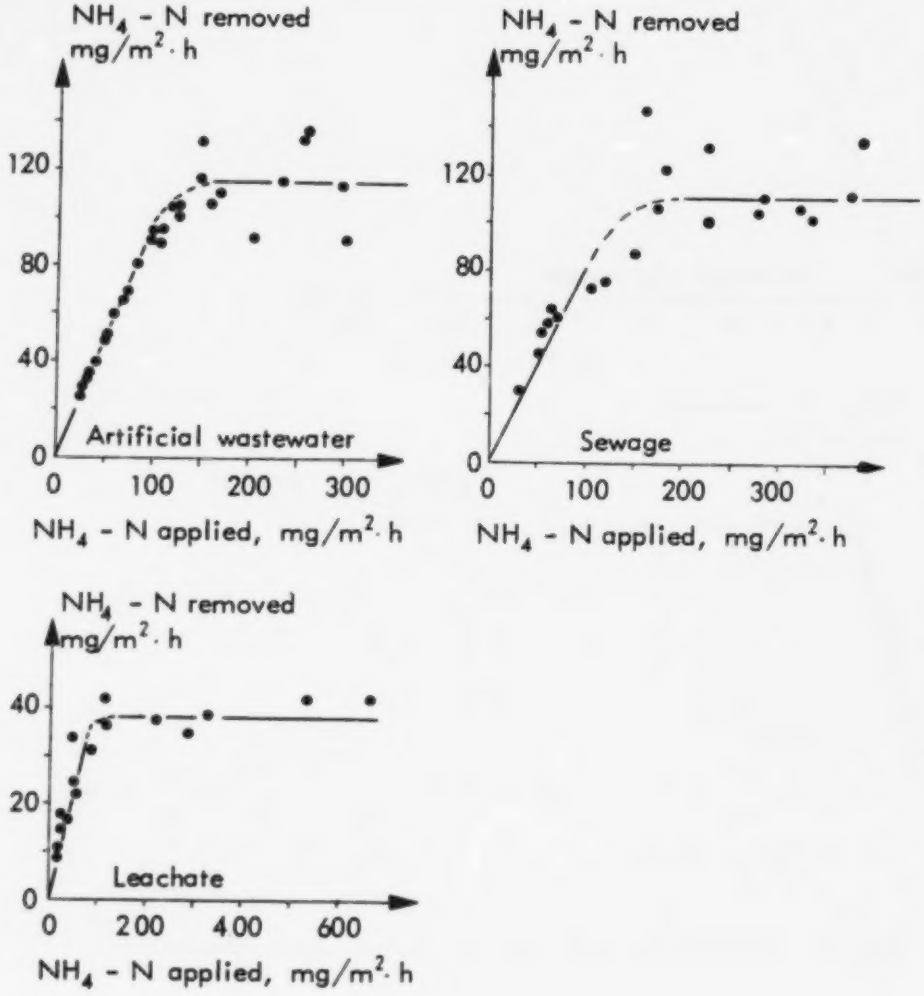


Fig. 7. Nitrification rate against ammonium-loading.

In Fig. 8 is shown the nitrification rate in each step of the unit, which may be assumed to be complete mixed.

Even though the spread in the results are great, it may be concluded that the nitrification process follows Monod-kinetics, with zero-order kinetics at higher ammonium concentrations and first order at lower concentrations. It seems that zero order kinetics governs the process at concentrations higher than 4-6 mg NH_4-N/l in normal sewages and higher than 20 mg NH_4-N/l in landfill leachate.

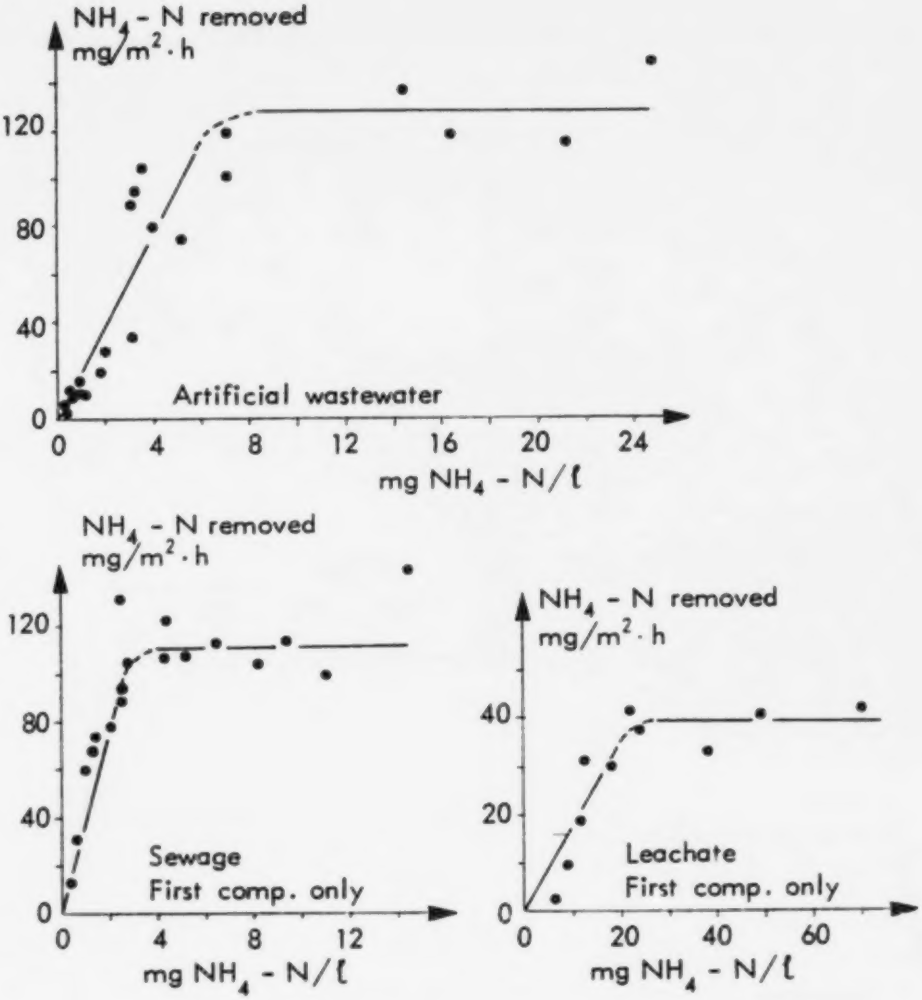


Fig. 8. Nitrification rate against ammonium-concentration.

In Table 3 the nitrification rates found in this study is compared to what has been found in other studies.

Table 3 Nitrification rates

Raw water	Nitrification rate	Reference
Sewage with $NH_4(HCO_3)$ addition	77.5 mg NH_4-N/m^2·h	Pretorius /10/
Municipal sewage	44 mg TKN/m^2·h	Murphy et al./11/
Artificial wastewater	max. 120 mg NH_4-N/m^2·h	This study
Municipal sewage	max. 110 mg NH_4-N/m^2·h	" "
Landfill leachate	max. 40 mg NH_4-N/m^2·h	" "

Denitrification

Several factors are influencing denitrification, such as oxygen concentration, pH, temperature, carbon source, nitrate-loading and ratio between loading of organic matter and nitrate.

The denitrification has to be carried out at anoxic conditions. True anaerobic conditions in the liquid does not seem to be necessary /12,4/. In the experiments presented here, conditions were normally anaerobic but denitrification took place even at oxygen-concentrations in the water around 1 mg O_2/l. It was not necessary to cover the free water surface to the atmosphere.

Optimum pH for denitrification seems to be pH = 7 - 7.5 /8/. In our experiments the pH varies between 7 and 8 in the wastewater experiments and between 8 and 8.5 in the leachate experiments.

Methanol is normally used as carbon source. Several authors have investigated nitrate/nitrite-removal as a function of methanol consumption for fixed film systems. Christensen and Harremoes /12/ reviewed the different investigations, and concluded that empirically, a methanol to nitrogen ratio of 3 g CH_3OH/g NO_3-N was necessary to reach more than 90% nitrate removal.

In suspended cultures the theoretical methanol consumption would be 2.47 g CH_3OH/g NO_3-N /13/, which is close to what has been measured by many authors /12/.

The most interesting alternative to methanol is undoubtedly the raw sewage itself. Narkis, Rebhun and Sheindorf /14/ have recently published results from suspended culture experiments where different carbon sources (methanol, sodium acetate and chemically treated raw sewage) were used. They concluded that by expressing the organic matter as soluble BOD, a critical value of 2.3 mg BOD/mg NO_x-N existed when 100% denitrification was to be reached regardless of what carbon source was used.

In the experiments presented here, soluble COD was, for different reasons, used as a measure for the organic matter. In the artificial wastewater experiments the COD-concentration did not seem to be limiting for denitrification. In the municipal wastewater experiments, sewage from morning hours was used. This had an abnormal high nitrogen concentration compared to the soluble COD-concentration. Complete denitrification was consequently not obtained. In Fig. 9 the nitrate removal is plotted against the COD/NO_x-N ratio (COD in raw water and total NO_x-N applied).

It is demonstrated that almost complete denitrification was obtained when the ratio of soluble COD in the raw water to the NO_x-N concentration into the anaerobic unit was over 7, and that the relationship between the removal and this ratio was linear for ratios lower than 7.

This made us add dried milk corresponding to a soluble COD addition of 100 mg COD_S/l in the proceeding experiments. (Bio-disc systems are, by the way, normally used in Norway when dairy effluents are treated together with municipal sewage.)

The fact that the influent COD is reduced through the anoxic step is another valuable aspect about the proposed process. Reduced organic loading on the aerobic step makes it possible to reduce the necessary area in this step in order to achieve the required COD to obtain nitrification. Altogether, this means lower investment cost compared to a situation were the denitrifier is proceeding the nitrifier.

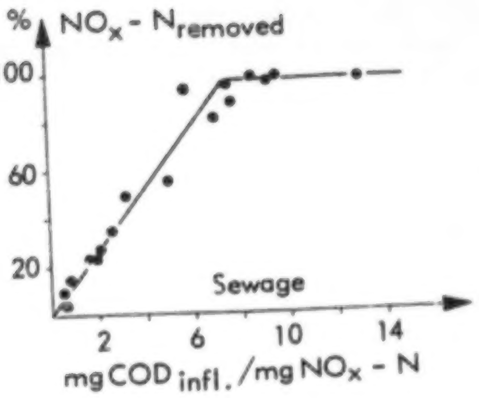


Fig. 9. Degree of denitrification against COD/NO_x-N ratio.

In order to find how much COD that was consumed by denitrification, the ratio between $COD_{consumed}$ and $(NO_x\text{-}N)_{consumed}$ was calculated. This ratio should be expected to be constant for the different runs. In order to visualize the results the ratio is plotted against recirculation ratio. Even if there is a considerable spread in the results, Fig. 10 shows that 2.5 - 3.5 mg COD is consumed for each mg NO_x-N removed by denitrification.

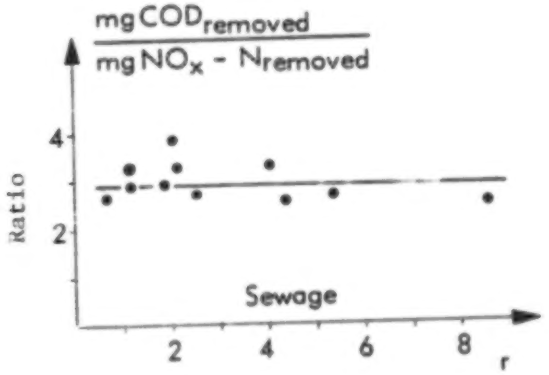


Fig. 10. COD consumed during denitrification.

In Fig. 11 are plotted the NO_x-N removal (%) versus NO_x-N loading for the different waters.

In the artificial wastewater experiment, a maximum load of 4 g NO_x-N/m^2·d could be tolerated in order to achieve more than 90% denitrification. The corresponding load in the sewage experiment seemed to be a little bit higher, 6 g NO_x-N/m^2·d.

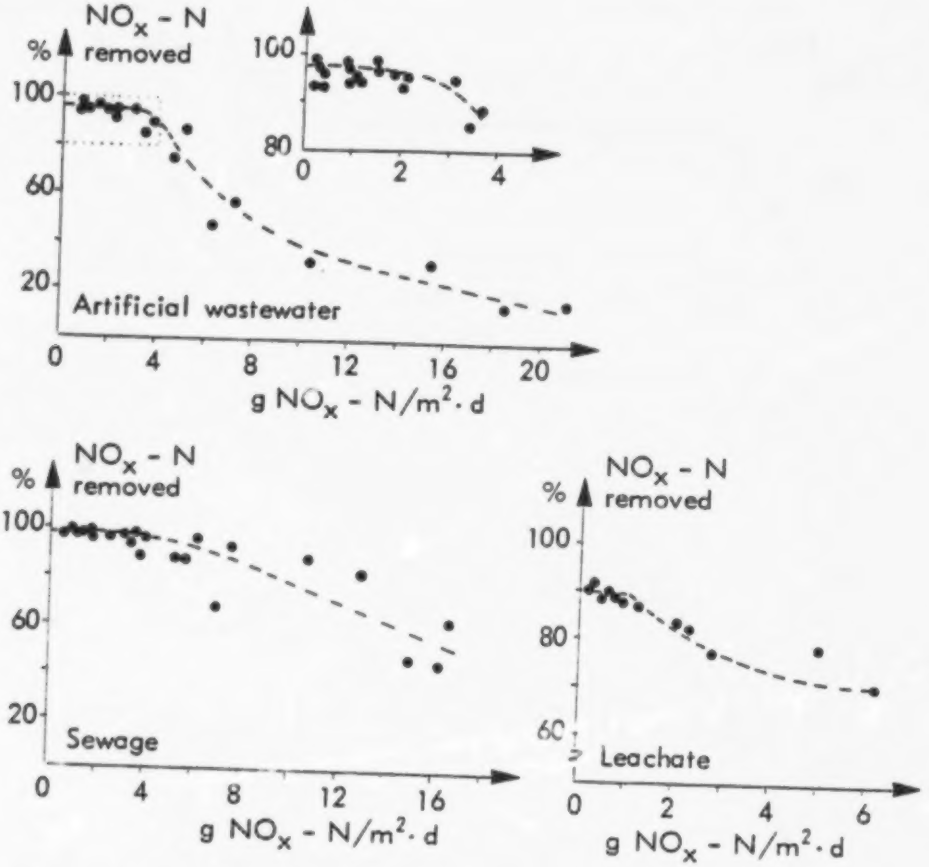


Fig. 11. Degree of denitrification against NO_x - N - loading.

In the leachate experiment more than 90% denitrification was difficult to achieve. This might be caused by some unknown limitation. To achieve 80% denitrification a maximum load of 2 g NO_x-N/m²·d could be used.

In Fig. 12 is plotted the denitrification rate (mg NO_x-$N_{removed}$/m²·h) against the NO_x-N load (mg NOx-$N_{applied}$/m²·h).

It is demonstrated that denitrification rate increases linearly with the load. In the sewage and the leachate experiments it is difficult to state what is maximum rate. In the sewage experiment it seems to be in the range of 350 - 450 mg NO_x-N/m²·h.

In the artificial wastewater experiment the maximum value is more pronounced at a value of 150 mg NO_x-N/m²·h.

The reason for this discrepancy is difficult to state, but we believe that the denitrification rate in the artificial wastewater experiments was limited by nitrate concentration at high denitrification rates.

Watanabe and Ishiguro /3/ found no maximum denitrification rate even if they measured rates as high as 694 mg NO_3-N/m²·h, while Cheung /4/ reported maximum denitrification rate of 676 mg NO_x-N/m²·h.

The denitrification rates found in this experiments are compared to what is found in other experiments in Table 4.

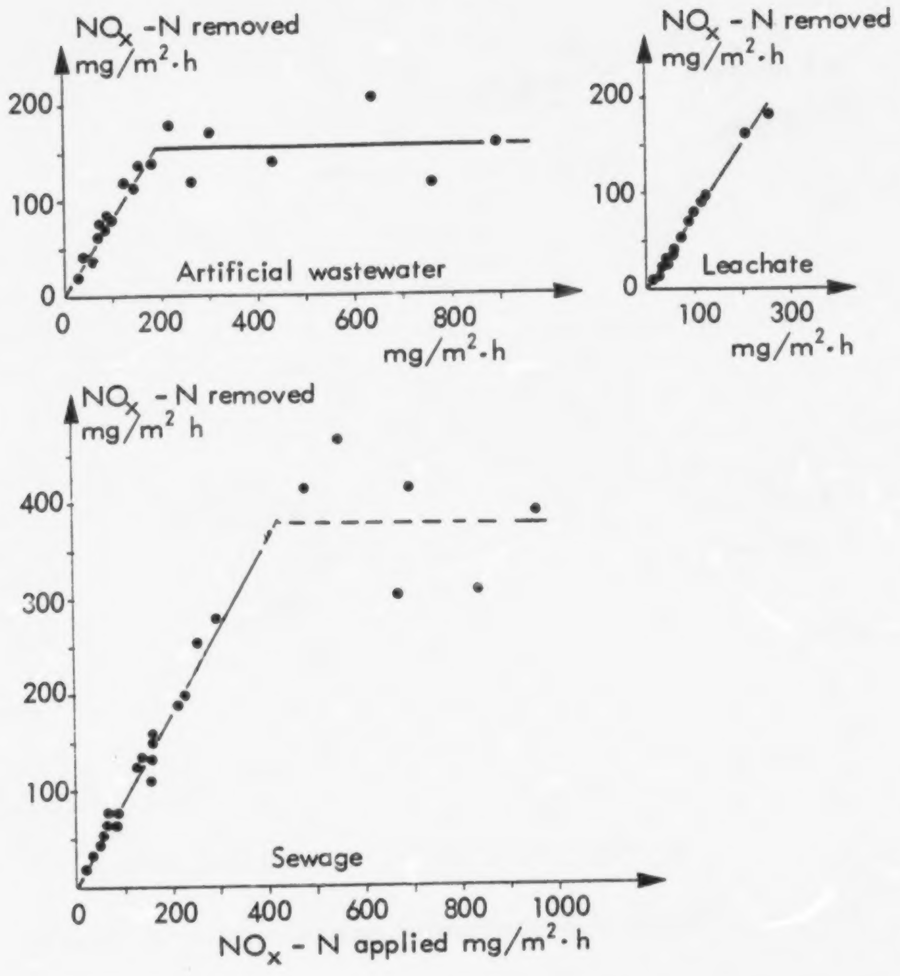


Fig. 12. Denitrification rate against NO_x - N loading.

Table 4 Denitrification rates

Raw water	Carbon source	Denitrification rate	Reference
Trickling filter effluent with KNO_3-addition	Methanol	max.260 mg NO_3-N/m²·h	Davies and Pretorius /1/
Effluent from biological treatment with $NaNO_3$-addition	Methanol	max.676 mg NO_x-N/m²·h	Cheung /4/
Nitrified sewage with KNO_3-addition	Methanol	1) 694 mg NO_3-N/m²·h	Watanabe and Ishiguro /3/
Artificial wastewater	Raw water	max.150-200 mg NO_x-N/m²·h	This study
Sewage	Raw water	max.350-450 mg NO_x-N/m²·h	" "
Landfill leachate	Raw water	1) 150-200 mg NO_x-N/m²·h	" "

1) Not maximum level.

DISCUSSION

The results of these experiments show that the proposed process should be an interesting alternative when nitrogen is to be removed from wastewater.

The advantages of the process are:

1. Low operation cost since external carbon source is omitted
2. Low capital cost since only one sedimentation unit is needed, and since the aerobic biodisc can be reduced as a consequence of organic matter being consumed during denitrification in the anaerobic step.
3. No danger of destroying the effluent BOD as a consequence of excessive methanol addition.

The possible disadvantages are:

1. The ratio between soluble COD loading (g COD_S/h) on the denitrifier from the influent and the NO_X-N loading (g NO_X-N/h) on the denitrifier should be ≥ 7.
2. High total nitrogen removals (>90%) requires a high recirculation ratio ($Q_R/Q > 9$), 80% removal is however reached at a Q_R/Q of 4. Recirculation may however be arranged very cheaply by a bucket-type pump mounted on the biodisc shaft.
3. If nitrification for some reason is inhibited, the ammonium loading on the nitrifier increases rapidly as a consequence of recirculation. This accelerates the reduction in nitrification. If there are signs that nitrification has been reduced, the recirculation ratio should be lowered until the nitrification is brought back to the normal state before recirculation is brought back to normal.

When designing a plant with the proposed process, three factors are to be determined, the recirculation ratio, the biodisc area for the nitrifier and the biodisc area for the denitrifier.

The recirculation ratio is determined by the percentage of nitrogen removal that is needed after the equation:

$$R_N = \frac{r}{r+1} \cdot 100\%$$

One should be conservative and use a bit higher recirculation ratio than that is found by this equation to account for the possibility that nitrification is not 100%.

The area of the aerobic nitrifier should be designed after normal design criteria in order to obtain nitrification on biodiscs. The organic load, however, may be reduced by 2.5 g COD for each g NO_X-N that is to be removed in the denitrifier. In normal sewage this would mean a reduction in organic load corresponding to a reduction in the concentration of the influent of 70 - 80 g COD/m^3 or 50 - 60 g BOD_5/m^3 (soluble organic matter). It is, however, important that the nitrifier is conservatively designed in order to ensure full nitrification. There is no need to add anything to the organic load from the recirculated water because this water has a very low concentration of biodegradable organic matter since the water already has passed the nitrification step.

The area of the denitrifier should be designed for a NO_X-N load of about 4 g NO_X-N/m^2·d to ensure full denitrification.

When these guidelines are followed, one would normally find that the necessary area for the denitrifier is less than half of the area of the nitrifier.

CONCLUSIONS

Nitrification/denitrification experiments with an anaerobic-aerobic biodisc system with recirculation have been carried out using artificial wastewater, sewage and landfill leachate as raw water. The following conclusions can be made:

1. It has been demonstrated that the proposed process (Fig. 2) are able to give nitrogen removals in sewage corresponding to $R_N = \{r/(r+1)\} \cdot 100\%$ (r = recirculation ratio) without the use of an external carbon source.
2. Nitrification rates of 120, 110 and 40 mg NH_4-N/m^2·h have been measured for the artificial wastewater, the sewage and the landfill leachate respectively.
3. Denitrification rates of 150-200, 350-450 and 150-200 mg NO_X-N/m^2·h have been measured for the artificial wastewater, the sewage and the landfill leachate respectively.
4. It has been found that the consumption of organic matter during denitrification is near 3 g COD/g NO_X-N removed.
5. In order to achieve full denitrification by the process, it has been found that the ratio between soluble COD-loading (g COD_S/h) on the denitrifier from the influent and the NO_X-N loading (g NO_X-N/h) on the denitrifier should be ≥ 7. This corresponds to a COD_S/N-ratio of 6-7 in the raw water.
6. In addition to the possibility of the process to treat sewage, the experiments also demonstrated that landfill leachate can be treated by the process. It is, however, recommended that iron in this type of water is removed prior to the biodisc process in order to avoid ironhydroxide precipitation on the biofilm.
7. The main advantage of the process is that addition of external carbon source is omitted and that the process would be economically favourable both from an investment and an operation viewpoint.

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NITRIFICATION/DENITRIFICATION STUDIES WITH ROTATING BIOLOGICAL CONTACTORS

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1.0 Introduction

Considerable research has been conducted over the past ten years on the use of biological methods for the removal of nitrogen from wastewater. Variations of the activated sludge (A/S) process have been utilized to first achieve nitrification (oxidation of ammonia to nitrate), then denitrification (the reduction of nitrate to nitrogen gas).

Nitrification/denitrification systems using the activated sludge principal consumes considerable energy and demands close operator attention.

With the current demands for nitrification by the Ministry of Environment (MOE) in some areas of southern Ontario (1) an alternative to the activated sludge system has been sought. The Rotating Biological Contactor (RBC) offers less energy consumption and operator attention than A/S systems and studies have shown comparable consistuent removals (2). As a result, following pilot-studies, a full-scale RBC section has been installed for nitrification of secondary effluent at the Guelph Ontario WPCP.

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The requirements for nitrification in Ontario are all site-specific. Generally, the concerns are the nitrogenous oxygen demand upon the receiving stream (mainly river systems) and the toxic effect of ammonia (in the unionized form) upon aquatic life.

There are some denitrification facilities treating high nitrate industrial wastewater, but to the author's knowledge, only one municipal plant (i.e. Penticton, B.C.) is currently being planned in Canada. Nitrogen removal from domestic wastewater is at the moment of low priority, trailing recent increased emphasis on better phosphorus removal.

Currently all of the municipal treatment plants expanded for nitrogen control in Ontario, will only include nitrification. If a step is required to reduce effluent nitrate (again site specific), for additional nutrient removal, a tertiary denitrification system run on nitrified secondary effluent would be a logical choice.

1.1 Past Studies

Investigations into the application of the RBC for tertiary denitrification have not been widely reported, but there are a few worth noting.

In 1974 Davies and Pretorius (3) published work done with an RBC unit converted for denitrification. By sealing the top of the unit, oxygen transfer to the discs was reduced creating an anoxic atmosphere conclusive to denitrification. The discs were rotated at 6 rpm and submerged to 45% of their area.

Parameters such as pH, temperature and methanol addition were examined as to possible effects on RBC denitrification efficiency. An optimum pH range of 7 to 8.5 for denitrification was found with 80% of maximum nitrate removal obtained at pH 6, and 70% at pH 9. An optimum temperature range of 10-30°C was reported and a methanol to nitrate (C/N) ratio of 2.6/1 was found adequate for complete denitrification.

Davies and Pretorius (3) commented, that a larger volume of liquid could be treated with the discs fully submerged, but corrosion of bearings can be a problem. Partially submerged discs would have more unit airspace and allow better release of nitrogen gas from the discs bio-growth.

More recent studies by Soyupak (4) (Master of Engineering Thesis), covered areas such as the effects of hydraulics and temperature on denitrification reactions within a totally submergenged disc unit. Pilot-plant size units with disc areas of $28.4 m^2$ ($306 ft^2$) were tested at flow-rates of 2.5 L/min. (0.6 Igpm) and 4 L/min. (0.9 Igpm) with the units operated single-stage or 4 stages in series. Hydraulic responses with the addition of dye at 4 different temperature levels were monitored after establishing biological growth on the discs.

Methanol and influent nitrate concentration were held constant at a C/N ratio of 1.5/1.

Out of this work, the results showed that each compartment (stage) of the RBC reacted (hydraulically) similiar to two Continuously Stirred Tank Reactors (CSTR) in series with interchanging flow.

Denitrification reaction rates were found to be zero order for the level of influent oxidized nitrogen in these studies.

The temperature dependency for denitrification using the RBC appeared not to be influenced by the hydraulic loadings studied. However denitrification rates were more sensitive to temperature variations than a CSTR unit or packed column reactors under similiar operating conditions.

Comparisons made on a volumetric bases, showed that the RBC produced higher denitrification rates than CSTR with a mixed liquor suspended solids (MLSS) range between 1500 and 2500 mg/L, or a packed column.

Pilot-plant studies by Antonie published in a text book on RBC's (5), indicated the possible nitrate removal efficiencics attainable at various hydraulic loadings. For example, using fully submerged discs at a wastewater temperature of $\geq 13^{\circ}C$ ($55^{\circ}F$) and nitrate concentrations up to 25 mg/L, 90% nitrate removal were predicted at a hydraulic rate of 0.15 $m^3/m^2/d$ (3 gpd/ft^2). Reduced nitrate removals down to 50% can be expected when increasing the hydraulic loading to 0.49 $m^3/m^2/d$ (10 gpd/ft^2) (See Figure 1). Methanol addition was implemented in these experiments, but the amount was not stated. However, the author did suggest that a floating cover could be placed on the water surface over the discs, to restrict oxygen transfer, thereby reducing methanol requirements.

After reviewing the literature it was felt that more specific information was required; particularly on a combined nitrification-denitrification RBC system.

2.0 Objectives of Study

A pilot-study was formulated and implemented to gain first-hand information as to the viability of an RBC system for nitrogen removal. The study was conducted at the Ministry of Environment Ontario Experimental Facility (O.E.F.) in Brampton.

The particular objectives of this project were to study the effects of:

(a) a nitrification RBC module operation on ultimate performance of a denitrification RBC module.

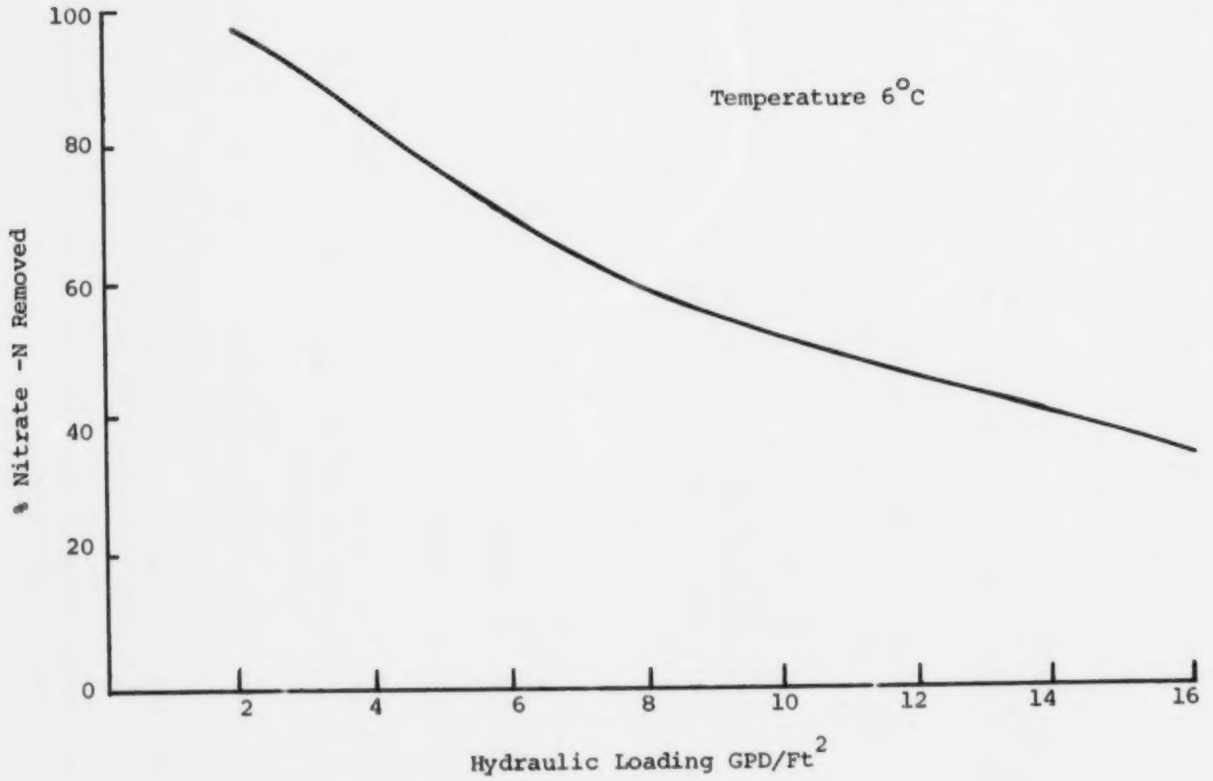
(b) submergence of the denitrification module discs on nitrate removal efficiency.

FIGURE 1

Nitrate Removal Vs Denitrification RBC

Hydraulic Loading

(reproduced after Antonie (5))



3.0 Experimental Setup And Procedures

3.1 Pilot-Plant Description

The pilot-plant consisted of two complete RBC modules arranged in series (Figure 2). The first module for nitrification had a total disc area of 156 m^2 (1680 ft^2) and comprised of four equally sized stages (compartments). The plastic discs were driven by an electrical variable speed motor, capable of delivering up to 9 rpm.

A portion of un-nitrified secondary effluent from a fullscale activated sludge process at O.E.F., was used as feed for the nitrification module. Previous studies by Hewitt of this Ministry (6) showed that 0.12 $m^3/m^2/d$ (2.4 $Igpd/ft^2$) would produce full nitrification during winter operation. An hydraulic loading of 10 L/min. (2.2 Igpm) approximating this value, for the nitrification RBC module, was selected for this study. The module was fed to give a plug-flow (4 stages in series) mode of operation.

The second disc module for denitrification had a total disc area of 23 m^2 (250 ft^2) and received a portion of nitrified effluent, from the first RBC module at feedrates between 2.5 and 5.2 L/min. (0.56 to 1.2 Igpm). These flowrates produced a hydraulic to surface loading range of between 0.16 to 0.34 $m^3/m^2/d$ (3.2 to 7 $Igpd/ft^2$) to a total of 4 stages in series.

Prior to the study, alterations were made to the module effluent piping to facilitate control of liquid level; from 45% to full disc submergence. For the same purpose, the walls of the module were also extended, the disc drive-shaft lengthened, and new bearings with snuffing boxes were installed to prevent leaks. The disc rotation speed was fixed at 6 rpm.

Sampling taps were installed along the base of the module, one at each RBC stage. A clear plastic cover was also placed over the top of the unit to reduce oxygen transfer to the discs, but to allow visual inspection.

Methanol was pumped to the first stage of the denitrification module at a predetermined concentration ratio to influent nitrate -N (C/N ratio) of 3 to 1. This ratio was calculated using a formula developed by McCarty (7) which is stated as follows:

$$Cm = 2.47\ N_o + 1.53\ N_i + 0.87\ D_o$$

where C_m = mg/L methanol
N_o = NO_3-N concentration mg/L
N_i = NO_2-N concentration mg/L
D_o = influent dissolved oxygen concentration mg/L

An influent nitrate -N of 15 mg/L, nitrite -N of 1 mg/L and DO of 8 mg/L were assumed for this study. Using these values in the above

FIGURE 2

Diagram of the O.E.F.

Nitrification/Denitrification RBC Pilot-Plant

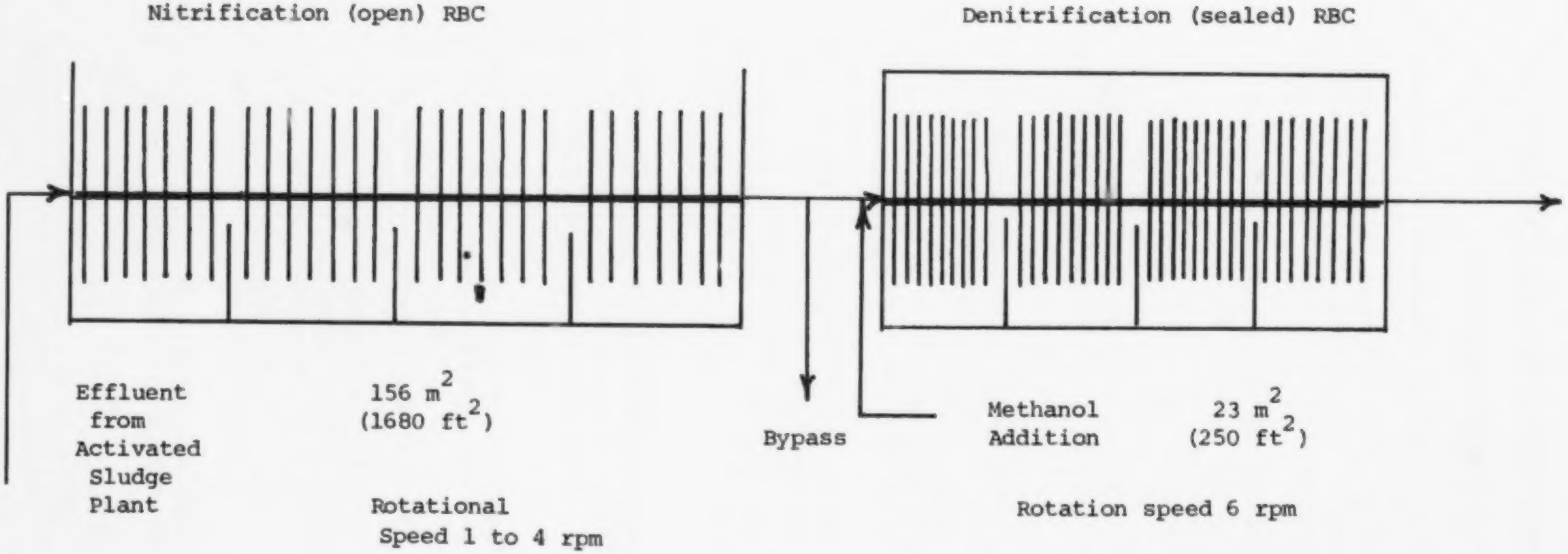

Nitrification (open) RBC
Denitrification (sealed) RBC
Effluent from Activated Sludge Plant
156 m^2 (1680 ft^2)
Rotational Speed 1 to 4 rpm
Bypass
Methanol Addition
23 m^2 (250 ft^2)
Rotation speed 6 rpm

TABLE 1

OEF RBC STUDY SCHEDULE

DENITRIFICATION

RUN #	DATE	W.W.T. °C	FEEDRATE L/MIN.	MODE	C/N RATIO
1	Oct. 2 26/78	19	3	Sub.	2.3/1
2	Nov.1-Dec.15/78	17	5	Sub.	2.0/1
3	Jan. 2-Feb. 9/79	11	3	Semi-Sub	1.6/1
4	Feb. 12-19/79	7	5	Semi-Sub	1.9/1
5	Feb. 19-Mar. 19/79	7	5	Semi-Sub	2.6/1
6	Apr. 19-May 2/79	11	5	Semi-Sub	3.2/1
7	May 7-June 8/79	14	5	Sub	3.7/1
8	June 11-June 22/79	15	5	Semi-Sub	3.4/1
9	June 25-July 11/79	16	2.5	Semi-Sub	3.4/1

* Specialized Run involving variation of nitrification module rpm through a total of 5 individual tests.

equation a methanol dosage of 45 mg/L (C/N of 3/1) to the influent flow, was required.

The nitrification and denitrification RBC modules were housed in an insulated trailer heated by a small electrical heater, to prevent freezing of effluent lines and samples.

3.2 Study Schedule

A series of experimental runs were conducted from October 1978 to July 1979. A total of 9 runs were completed in which the nitrate removal efficiency of the denitrification RBC module was compared under semi-and total disc submergence, through naturally varying wastewater temperatures.

Disc rotational speed in the nitrification module was fixed at 4 rpm for runs 1 to 4 and 2 rpm in runs 6 to 9.

More detailed investigation was given to the nitrification RBC module during run #5. The module was operated at 1 to 4 rpm to determine the effects of disc rotational speed on module DO and resultant nitrification. The effects of the last stage DO (nitrification module) level on nitrate removal in the denitrification module was also studied.

Half-way through the study, the methanol stock was found to be 20% lower concentration then expected. Consequently, for the first four runs methanol addition was 1/3 lower-than the requirement.

A schedule showing modes of operation for the RBC study and corrected C/N ratio is contained in Table 1. The study was primarily formulated on the operation of the denitrification RBC module.

3.3 Analyses And Tests

Twenty-four hour composite samples were taken (min. 3 days/wk.) of nitrification module influent and effluent and denitrification effluent. Allyl-thiourea (ATU) was added to each sample container to a concentration of 1 mg/L, as a preservative for nitrification.

The following analyses were performed:

- Five-day Biological Oxygen Demand (BOD_5)
- Suspended Solids (SS)
- Total Kjeldahl, ammonia, nitrite and nitrate nitrogens (TKN, NH_4-N, NO_2-N and NO_3-N)
- Chemical Oxygen Demand (COD)
- Alkalinity as $CaCO_3$

Onsite tests routinely consisted of stage by stage measurement of DO and temperature twice per day. These measurements were taken more frequently during intensified work in run 5, and continued into runs 6 to 9.

Dissolved oxygen measurements on the denitrification module were conducted, by first evacuating a 1 Litre flask(containing a DO probe) with helium, and then introducing wastewater from the sampling taps on the side of the unit.

4.0 Discussion of Results

4.1 Nitrification RBC Module

(a) Runs 1 to 5

For the most part, complete oxidation of ammonia was accomplished in the nitrification module, producing a mean effluent ammonia of < 1 mg/L, at a mean influent ammonia concentration of 14 mg/L. (See Table 2). The resulting range of hydraulic loading to the nitrification module for the study was 0.076 to 0.102 $m^3/m^2/d$ (1.5 to 2.0 $Igpd/ft^2$).

Exceptions to this were shown in runs 3 and 5. In run 3, the effluent ammonia concentration rose slightly to approximately 1.3 mg/L and coincided with a 6°C drop in temperature. The highest influent ammonia to disc surface loading (1.68 $gNH_4-N/m^2/d$) also occurred during this run (Table 2).

As stated earlier, more intensive work was done on the nitrification module in run 5. Routine DO measurements (during runs 1 to 4) in the nitrification module showed levels often exceeding 8 mg/L and at times as high as 11 mg/L in the last two stages. These DO levels were higher than expected, and in turn introduced more demand for organic carbon, then orginally accounted for.

Consequently run 5 was formulated to see what effects various disc rpm had on nitrification efficiency, module DO level and ultimate nitrate removal in the second module.

Run 5 produced the highest mean effluent ammonia (2.1 mg/L) with an approximate 10% drop in removal efficiency resulting from reduced disc rpm. Surveys showed a marked difference in DO concentration in various RBC stages upon altering the disc rpm. A mean RBC DO of 3.6 mg/L resulted at 1 rpm compared to 8.7 mg/L at 4 rpm (See Table 3).

Ammonia oxidation, at liquid temperatures ranging from 6.5 to to 8.7°C, improved as the disc rpm was raised. Effluent ammonia -N ranged between 2.5 and 2.7 mg/L at 1 and 2 rpm, but fell below 1 mg/L at 3 and 4 rpm.

TABLE 2

NITRIFICATION RBC MODULE

ANALYTICAL RESULTS AND REMOVAL RATES

Run #	INFLUENT		Ammonia -N mg/L		NH_4-N Removal
	Temp °C	g NH_4-N/m^2/d	Influent	Effluent	g NH_4-N rem./m^2/d
1	19	0.98	11.4	0.24	0.69
2	17	1.60	16.8	0.60	1.12
3	11	1.68	17.1	1.27	1.13
4	7	1.28	17.1	0.68	0.89
5	7	1.09	15.0	2.10	0.69
6	11	0.98	11.5	0.84	0.67
7	14	0.58	6.7	0.20	0.41
8	15	1.24	14.0	0.73	0.86
9	16	1.22	15.8	0.68	0.85

TABLE 3

SUMMARY OF NITRIFICATION RBC DO PATTERNS AND EFFICIENCY

(Mean Values) Run 5

Disc RPM And Test #

	1	2		3	4
		(a)	(b)		
Flow 1/min	7.9	8.3	9.2	8.2	7.3
Igpm	1.85	1.84	2.04	1.82	1.62
Igpd/ft^2	1.54	1.53	1.70	1.52	1.35
Temp $^\circ$C	8.1	7.6	7.1	6.5	8.7
RBC DO					
Stage #1	4.4	6.9	8.9	8.6	8.8
(mg/L) #2	3.2	5.5	7.6	7.8	8.0
#3	3.3	5.2	7.1	8.5	8.1
#4	3.5	6.3	6.9	9.5	10.0
NH_4-N					
Influent g/d	205	191	172	154	168
Removal g/d	174	160	139	146	160
Removal %	85	84	81	95	95
Effl. con. mg/L	2.7	2.6	2.5	0.6	0.8

The tests at 2 rpm were carried out for a longer duration and divided into two sub-programs (a) and (b), to determine the effects of disc mass loading. At similiar rpm and temperature, little difference in nitrification was noted between a disc ammonia loading of 160 g/d and 139 g/d (Table 4).

(b) Runs 6 to 9

Following run 5, the disc rpm was maintained at 2 for the remainder of the study. It was felt that the expected rise in temperature during the coming spring and summer would compensate for any loss of nitrification at a lower disc rpm. The analytical results from runs 6 to 9 show consistant effluent ammonia of less-than 1 mg/L, at mean wastewater temperatures ranging from 11 to 16°C (Table 2). These results were achieved at a mean RBC DO ranging from 4.3 to 7.3 mg/L. The effect of reduced oxygen demand was shown in run 7 (Table 2) when the influent ammonia concentration dropped to 6.7 mg/L, coincidental with a low BOD_5 of 9 mg/L. As a result, the mean RBC DO rose approximately 2.0 mg/L above that of runs 6, 8 and 9.

4.2 Denitrification RBC Module

(a) Run 1 to 4

As stated earlier, resultant C/N ratio in the first 4 runs was lower-than predicted because of a dilute methanol supply. For this reason the initial runs were only compared, on the basis of denitrification efficiency, with each other.

A drop in nitrate removal from 73% to 31% occurred when the feedrate of the denitrification module, at full disc submergence, was increased from 3 to 5 L/min. (Runs 1 and 2). On a weight to disc area basis, the nitrate loading rates for run 1 and 2 were 2.68 and 4.87 g NO_3-N/m^2/d respectively (Table 5).

A further drop in nitrate removal resulted in runs 3 and 4 when the module was changed to the semi-submerged mode. Values of 26% and 28% removal were produced in runs 3 and 4 respectively, at nitrate loadings slightly higher than the previous two runs (Table 5).

Taking into account a slightly higher nitrate loading in run 4 (compared to run 2), it appeared that both semi and submerged modes produced identical results at the higher loadings (4.87 and 5.49 g NO_3-N/m^2/d) at similiar C/N ratio. However better results were shown in the submerged mode at the lower (3 L/min) feedrate. Wastewater temperature dropped from a mean of 19°C to 7°C from runs 1 to 4, and probably was partially responsible for a loss in denitrification during the semi-submerged runs.

More definite comparisons resulted between submerged and semi-submerged modes of operation, in the latter part of the study at a narrower temperature range.

TABLE 4

RBC DO SURVEYS

NITRIFICATION

RUN #			MODULE STAGES 1	2	3	4	MEAN
6	MEAN		5.0	4.1	4.2	4.7	4.5
	RANGE	LO	4.2	3.8	3.8	4.2	
		HI	5.6	4.6	4.6	5.8	
7	MEAN		7.0	7.0	7.3	8.1	7.3
	RANGE	LO	3.5	3.4	3.9	5.0	
		HI	10.5	10.6	10.6	10.5	
8	MEAN		4.0	3.6	4.8	6.6	5.2
	RANGE	LO	3.6	2.9	4.0	5.6	
		HI	4.3	4.8	6.0	7.2	
9	MEAN		3.6	3.2	4.3	6.3	4.3
	RANGE	LO	3.3	2.7	3.9	6.1	
		HI	4.4	3.5	4.8	6.7	

Table 5

DENITRIFICATION RBC MODULE

LOADING AND REMOVALS (RUNS 1-4)

Run #	Mode	T^{o}C	DISC LOADING L/min	g Nitrate/m^{2}/d	Nitrate Removal %	g Nitrate/m^{2}/d	C/N Ratio
1	Sub	19	3.1	2.66	73	1.94	2.3/1
2	Sub	17	4.9	4.87	31	1.52	2.0/1
3	Semi -sub	11	3.0	2.94	25	0.76	1.6/1
4	Semi -sub	7	5.2	5.49	28	1.55	1.9/1

(b) Run 5

As a result of raised nitrification RBC DO (discussed earlier), DO concentration also rose in all stages of the denitrification module. As shown in Figure 3, values above 1 mg/L DO resulted in the denitrification module as influent DO (nitrification effluent) rose from 3.5 to 6.8 mg/L. A further increase to slightly more than 2 mg/L DO was observed in the denitrification module, when the nitrification module DO rose above 8 mg/L.

Little or no change was noted in nitrate removal between test 1 and 2(a) (test number also corresponds to nitrification disc rpm) at the same C/N ratio with a slight rise in module DO (Table 6). However a marked improvement did occur in test 2(a), when the C/N ratio was raised to 3.3, and illustrated the benefits of raising methanol dosage.

A drop in nitrate removal (68% to 18%) resulted in Test 3 at a similiar C/N ratio to test 2(b) when a rise in module DO occurred. Similiar results were shown in test 4 (24% nitrate removal) to previous runs 3 and 4 in the submerged mode (compared Tables 5 and 6).

It appeared from these tests, that a mean denitrification module DO of less than 2 mg/L, combined with a C/N ratio (3.3/1) close to the theoretical requirement (i.e. in test 2(b)), produced good nitrate removal. Consequently the nitrification RBC module rpm was lowered to 2, for the remainder of the study and the C/N ratio was maintained at approximately 3/1.

Runs 6 - 9

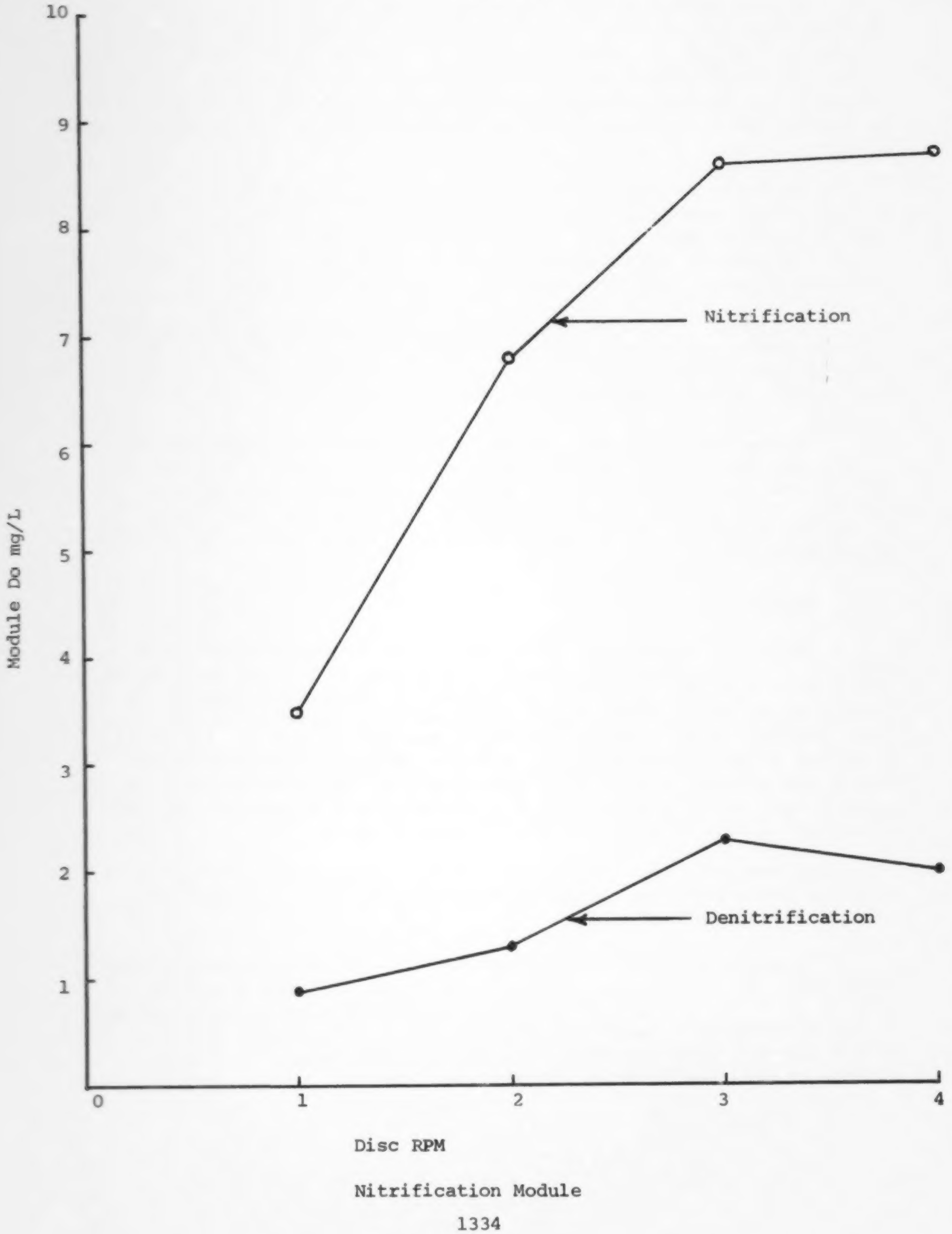
Under the revised operating conditions in runs 6 to 9 a wide range of nitrate removal efficiency took place, at C/N rations ranging from 2.8 to 3.2 (See Table 7). A wastewater temperature range of 12 to 16°C occurred during these runs. Similiar removals were observed in run 6 at identical module operation (semi-submerged) to run 5, (test 2(b)) at a liquid temperature of 11°C. However, when the denitrification module was changed to full disc submergence in run 7, an increase in nitrate removal resulted to 91% to a denitrification rate of 3.41 gm nitrate/m^2/d at a temperature of 14°C (See Table 7).

An immediate loss in nitrate removal efficiency to 40% occurred when the denitrification module was reverted to the semi-submerged mode (Run 8) at 16°C (Table 7).

In order to achieve better denitrification under semi-submerged conditions, the disc loading was reduced in the last run from 4.39 (run 8) to 2.27 g nitrate/m^2/d. An unexpected drop in nitrate removal to 23% occurred as a result of this change. A slight drop in C/N ratio to 2.8 may have contributed to this loss in efficiency, but higher nitrate removal was expected with reduced loading.

Effects of Nitrification RBC
RPM on Nitrification and
Denitrification Module DO Levels
(RUN 5)

FIGURE 3



Nitrification Module

TABLE 6

SUMMARY OF DENITRIFICATION RBC

DO PATTERNS AND EFFICIENCY

Nitrification Module Run 5

Disc RPM and Test #

	1	2		3	4
		(a)	(b)		
Flow l/min	5.6	5.2	5.5	5.5	5.6
NO_3-N					
Inf. g/d	137	120	107	119	137
Temp °C	9.0	8.6	7.3	7.0	9.6
C/N Ratio	2.0/1	2.0/1	3.3/1	3.0/1	2.6/1
RBC					
DO Bay #1 (mg/L)	1.3	1.2	1.8	2.8	2.2
#2	0.9	1.3	1.6	2.4	2.3
#3	1.1	1.6	1.1	2.2	1.6
#4	0.8	1.1	1.2	2.1	1.8
NO_3-N					
Removal $g/m^2/d$	1.59	1.55	3.14	0.91	1.42
" %	27	30	68	18	24
Effluent					
mg/l	12	11	4.4	12	13

TABLE 7

DENITRIFICATION RBC MODULE

LOADING AND REMOVALS

RUNS (6-9)

Run #	Mode	Disc Loading L/min	g Nitrate/m^2/d	C/N Ratio	%	Nitrate Removal g Nitrate/m^2/d	T°C
6	Semi-sub	5.0	4.36	3.0/1	64	2.76	12
7	Sub.	5.0	3.76	3.2/1	91	3.41	14
8	Semi-sub	4.5	4.39	3.2/1	40	1.70	16
9	Semi-sub	2.0	2.27	2.8/1	23	1.24	16

In the semi-submerged mode runs, a loss in denitrification efficiency resulted when the disc loading was reduced (compare runs 6 to 9). These results were coincidental with what occurred, less dramatically in runs 3 and 4.

DO surveys of the denitrification module through the last four runs showed mean values below 1.5, with the lowest at 0.8 mg/L in run 6 (Table 8).

DO levels between the module liquid surface and cover (module atmosphere) showed mean values between 5.1 and 8.7. The highest DO concentrations appeared in run 7 with full disc submergence and reduced air space over the module discs.

No correlation was found between RBC denitrification efficiency and DO concentration ranges occurring in the module during runs 6 to 9.

4.3 Removal of BOD_5, S.S. and TN (Total Nitrogen)

BOD_5 removal within the nitrification module, at 4 rpm (Runs 1 to 4), averaged 28% with influent levels of 25 mg/L and effluent values of 18 mg/L. Similiar values and reductions were observed for suspended solids (See Table 9).

Little change was noted in either BOD_5 or S.S. from influent to effluent in the denitrification module. A somewhat different pattern was shown in runs 5 to 9 at lower (2) disc rpm and influent constituent concentrations. A mean BOD_5 reduction from 12 to 7 mg/L resulted but S.S. rose from 8 to 11 mg/L. Therefore, soluble BOD_5 reduction through the nitrification module resulted during the latter runs.

A gain in S.S. through the denitrification module was shown; probably due to solids sloughing from the discs (Table 9). A long stringy loosely-attached growth was evident on the discs and bottom sediment was more pronounced than in the nitrification module.

For the most part, a slight gain in BOD_5 was noted in the denitrification module effluent, and was mainly attributed to increased suspended solids concentration.

4.4 Alkalinity

Previous studies by the author (8) on an activated sludge nitrification/denitrification system showed a wt./wt. ratio of calcium alkalinity removed per ammonia oxidized of 7.1. In terms of denitrification, a wt/wt. ratio of 4.2 of alkalinity returned to the system per nitrate removal was also observed.

In comparison, the RBC pilot-study showed calcium alkalinity loss ratio of 7.6 for nitrification and a gain ratio 3.7 for denitrification.

TABLE 8

RBC DO SURVEYS

DE-NITRIFICATION

RUN #			MODULE STAGES 1	2	3	4		MODULE ATMOS.*
	MEAN		1.0	0.9	0,6	0.7	0.8	5.1
6	RANGE	LO	0,8	0,7	0.5	0.4		4.9
		HI	1.4	1.3	0,8	0.9		5,2
	MEAN		2.1	1.5	1.2	0.9	1.4	8.7
7	RANGE	LO	1.3	0.8	0,8	0.6		2.7
		HI	3.0	2,4	1,7	1.5		10.8
	MEAN		1.7	1.5	1.2	1,1	1.4	6.6
8	RANGE	LO	1,2	1.0	0,9	0.8		5.2
		HI	2.5	2.5	2.4	1.5		10.0
	MEAN		1.5	1,2	1.3	1.1	1.3	5.7
9	RANGE	LO	1.3	1.0	0,8	0.8		4.9
		HI	1.7	1.6	1.8	1,2		6.4

* Airspace above disc

TABLE 9

RBC BOD_5, SS AND TKN RESULTS

(MEAN VALVES)

Nitrification				Denitrification	
Runs 1-4	Inf.	Eff.	% Rem	Eff.	% Rem
BOD	25	18	28	17	6
SS	21	15	29	16	- 7
TKN	20	3.5	82	3.9	-11
Runs 5-9					
BOD	12	7	42	10	-43
SS	8	11	-62	18	-36
TKN	15	3.4	77	3.7	- 9
Study Mean					
BOD			19		-19
SS			-17		-22
TKN			80		-10

5.0 Conclusions

The following conclusions were derived from the RBC pilot-plant study at O.E.F.

(a) Close to full ammonia oxidation was attained in the nitrification RBC module with some benefit shown on raising disc rpm above 2.

(b) DO concentration in the nitrification model increased with raised disc rpm and in turn lowered nitrate removal in the denitrification module.

(c) A nitrate removal efficiency of 91% with effluent nitrate level close to 1 mg/L, was attainable when the denitrification module was operated at full disc submergence. Lower nitrate removals were experienced with the semi-submerged mode at comparable liquid temperatures and C/N ratio.

(d) A reduction in BOD, both soluble and insoluble, resulted through the nitrification module with a slight gain in suspended matter shown in the denitrification model.

(e) Calcium alkalinity loss per weight ammonia oxidized (ratio) slightly higher than previous studies was shown; whereas alkalinity released per weight nitrate removal was less than previously observed in activated sludge process studies.

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PART XI: SELECTIONS AND ECONOMICS

DESIGN CONSIDERATIONS FOR A 16 MGD ROTATING BIOLOGICAL CONTACTOR TREATMENT FACILITY

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General

Eau Claire, Wisconsin is located approximately 100 miles east of the Twin Cities of Minneapolis - St. Paul, Minnesota. Eau Claire is one of the fastest growing metropolitan areas in the State of Wisconsin. The estimated 1980 population for the City is 50,000. The project population in the design year of 2000 is 101,716.

The climate in Eau Claire is one of extreme. Temperatures vary from above 95°F in the summer to below -30°F in the winter. Daily temperature variances of 30°F are common. The winter climate often is below 0°F for periods of two months or more. A typical winter will have snow covered ground for a three to four month period.

In 1969 the City took the initial steps to upgrade an overloaded primary treatment facility to provide secondary treatment of its wastewater prior to discharge to the Chippewa River, a tributary of the Mississippi. OWEN AYRES & ASSSOCIATES INC was hired as consulting engineer for the project. A preliminary engineering report completed in May, 1971 analyzed

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the treatment alternatives available to the City and recommended a treatment facility utilizing activated sludge treatment with solids incineration. By mid-1972 a design utilizing the report concept was approximately 60% complete.

With the passage of Public Law 92-500 in October, 1972 funding of the project became uncertain and design work was ceased. By late 1974 new funding regulations were being finalized and the City applied for funding to conduct a "201" facilities plan. The application was approved and work was restarted in early 1975 at the initial conceptual stages. The following numbers and costs are based upon the 1976 facilities plan and pilot studies conducted in 1976 and 1977.

Although frustrating to all parties involved, the delay in funding was beneficial. An opportunity was afforded for the analysis of newer technology and an analysis of more energy efficient systems was now feasible.

RECEIVING WATERS

The ultimate receiving water for the Eau Claire wastewater treatment plant is the Chippewa River. The River has its origin in a number of lakes and swamps in the northwestern part of the state. The drainage area is oriented in a northeast to southwest direction extending from near the Upper Michigan boundry to the confluence with the Mississippi River. The Chippewa drainage basin encompasses a total drainage area of 9468 square miles, being second in size in the State only to the Wisconsin River Basin. The total length of the basin is about 170 miles. The Eau Claire urban area is located about 55 miles from the Mississippi River, or in the lower third of the drainage basin. The Eau Claire River is a major tributary, its confluence with the Chippewa being in downtown Eau Claire.

Upstream from the urban area are approximately twenty-two municipalities, several milk processing plants and three paper mills that discharge from point sources various concentrations of wastewater to the Chippewa River or its tributaries. Many additional non-point wastewater sources exist upstream from Eau Claire.

STANDARDS AND GUIDELINES

The intrastate standards applicable to the Chippewa River in the Eau Claire area are those dealing with recreational use, fish, and other aquatic life. In summary, these standards are:

Dissolved Oxygen: The dissolved oxygen content in surface waters not be lowered to less than 5 mg/l at any time.

Temperature: There shall be no temperature changes that may adversely affect aquatic life. Natural daily and seasonal temperature fluctuations shall be maintained. The maximum temperature rise at the edge of the mixing zone above the existing natural temperature shall not exceed 5°F for streams. The temperature shall not exceed 89°F for warm water fish.

pH: shall be within the range of 6.0 to 9.0, with no change greater than 0.5 units outside the natural seasonal maximum and minimum.

Toxics: Unauthorized concentrations of substances are not permitted that alone or in combination with other materials present are toxic to fish or other aquatic life.

Bacteriological: The membrane filter fecal coliform count shall not exceed 200 per 100 ml as geometric mean based on not less than 5 samples per month, nor exceed 400 per 100 ml in more than 10% of all samples during the month.

BOD_5:	(monthly average)	30 mg/1
	(weekly average)	45 mg/1

Suspended Solids:	(monthly average)	30 mg/1
	(weekly average)	45 mg/1

The above standards are considered best practicable waste treatment technology to meet the state water quality standards. Any wastewater treatment scheme proposed must be capable of meeting these effluent standards as a minimum.

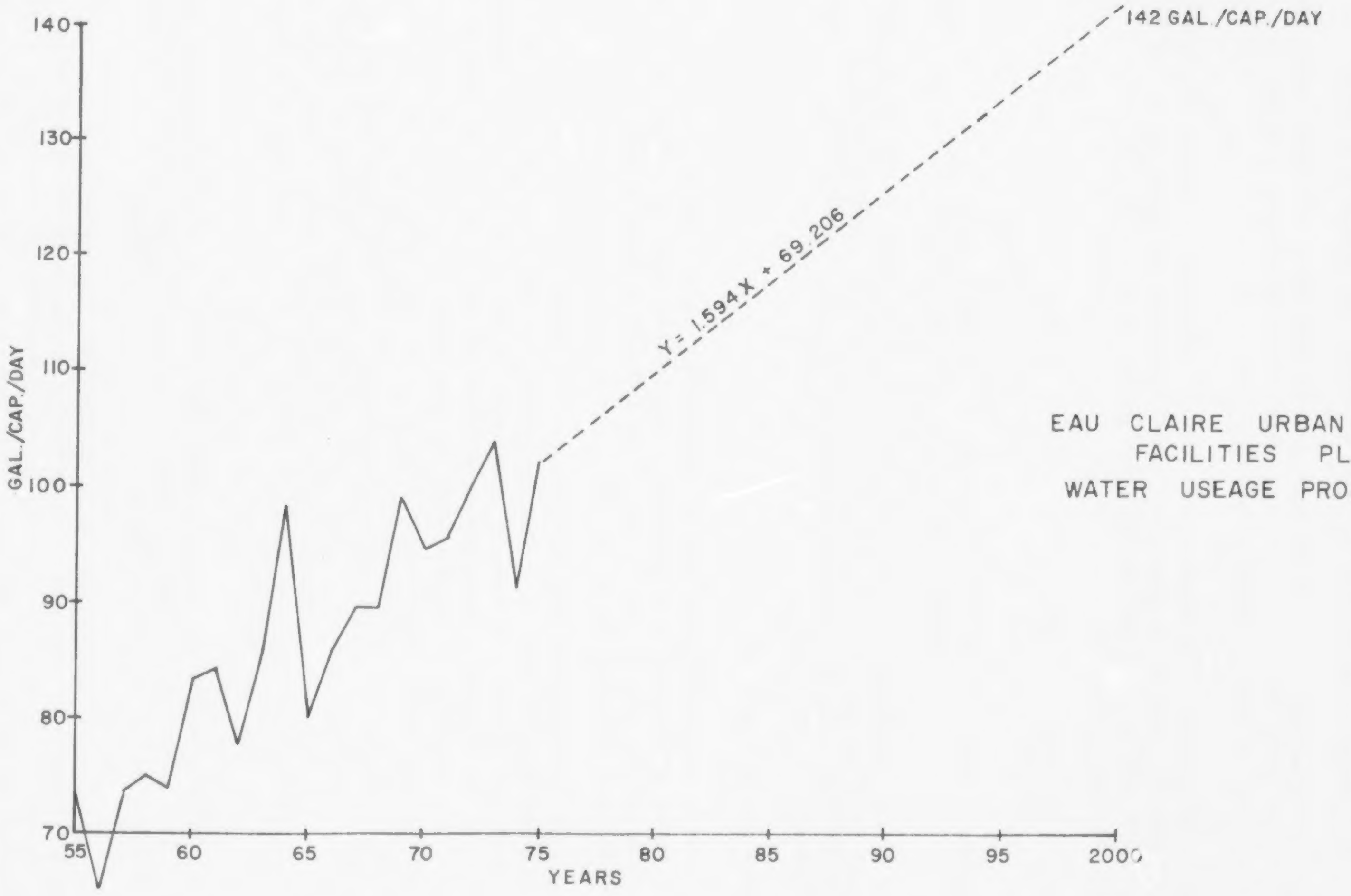
WASTEWATER PROJECTIONS

The three major contributors of wastewater in the Eau Claire urban area are those eminating from domestic, industrial and infiltration/inflow sources. A projection of the volume from the three sources was made for the service area through the year 2000.

Domestic wastewaters were defined as all flows from residential, commercial and public sources. The inaccuracy of the existing venturi flow meter at the Eau Claire wastewater treatment facility made use of this data questionable. To obtain a reasonable estimate of past and future wastewater flume, billing records available from the Eau Claire water utility were used. Per capita water consumption records were analyzed for the years 1955 through 1975. A linear regression analysis was used to predict water consumption in the year 2000 as shown by Figure 1. The year 2000 domestic water usage was projected to be 142 gallons/capita/day. Assuming 80% of the water usage eventually reaches the wastewater treatment plant, the per capita wastewater contribution will be 114 gpcd. The anticipated domestic wastewater flow will be 11.6 MGD.

Industrial wastewaters are the most difficult of all flows to accurately project. Two major industries are currently in operation within Eau Claire and discharge to the municipal facility wastewater system.

National Presto Industries is an ordinance plant which manufactures military projectiles. The market for the products from the plant vary widely. If the plant were to operate at full production the company estimates a wastewater volume of 608,000 gpd. This was used as a year 2000 design flow.



EAU CLAIRE URBAN AREA
FACILITIES PLAN
WATER USEAGE PROJECTION

FIGURE 1

OWEN AYRES & ASSOCIATES

Uniroyal, Inc. manufactures various types of rubber tires in Eau Claire. During a peak production day an estimated flow of 850,000 gpd is discharged into the sanitary sewer system. The wastewater is from wet collectors which trap carbon blank solids and from the cafeteria and restrooms. Uniroyal does not anticipate any substantial increase in production and a year 2000 flow of 850,000 gpd was used for design purposes.

Numerous other smaller industries are located within the City of Eau Claire. The average annual daily water usage for the industries is 0.9 MGD. The City is currently pursuing an aggressive program of obtaining small industrial facilities within the City. Based on industrial land available within the community an anticipated industrial wastewater flow from the smaller industries is 2.4 MGD in the year 2000.

An infiltration/inflow analysis prepared for the City of Eau Claire revealed excessive clearwater sources. After corrective measures are completed it is estimated that clearwater will be reduced to 0.8 MGD.

The anticipated year 2000 wastewater flows are summarized in Table 1.

YEAR 2000 WASTEWATER FLOWS
TABLE 1

Source	Volume (MGD)
Domestic Wastewater	11.6
Uniroyal, Inc.	0.85
National Presto Industries	0.6
Other Industry	2.4
Infiltration/Inflow	0.8
TOTAL	16.3

Other parameters of importance that required consideration were the 5-day biochemical oxygen demand (BOD_5) and the suspended solids concentration of the incoming wastewater. Past records available from area wastewater treatment facilities showed that a reasonable value of these parameters was 250 mg/l or 33,735 lb/day each.

LIQUID HANDLING

Only secondary treatment processes will be discussed in the section although a similar type analysis was conducted within other areas of the plant. Based on existing performance of the City treatment plant it was estimated that 35% of the incoming BOD_5 would be removed in the primary treatment portion of the facility. This leaves an organic loading (BOD_5) to the secondary process of 21,900 lbs/day.

Three secondary processes were analyzed in detail. These were an activated sludge process, rotating biological contactor (RBC) and an activated biological filter (ABF). Only the RBC process design will be detailed here.

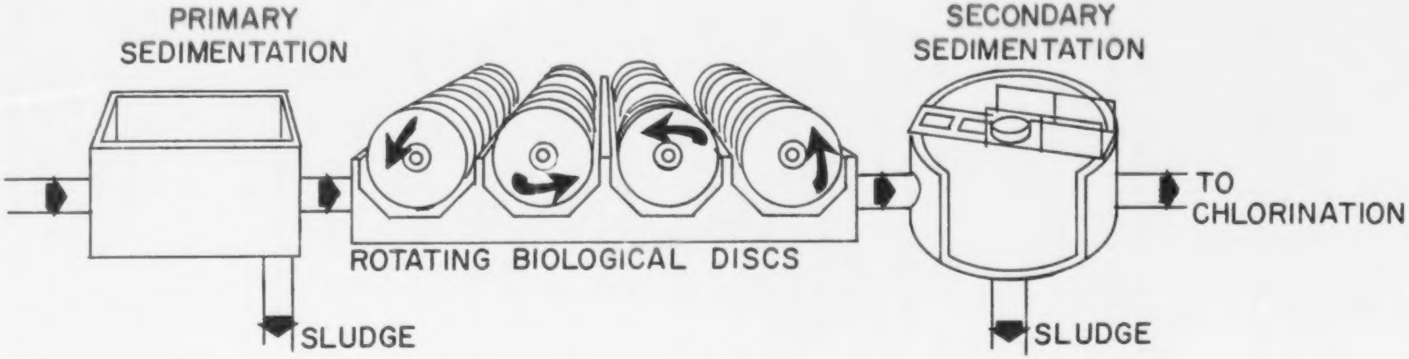
A schematic diagram of the RBC system is shown on Figure 2. The process consists of large diameter corrugated plastic media which is mounted on a horizontal shaft and placed in a concrete tank. The media is slowly rotated while approximately 40% of the surface area is submerged in the wastewater. Microorganisms naturally present in the wastewater adhere to the rotating surfaces. In rotation, the media carries a film of wastewater into the air where it trickles down the surface and absorbs oxygen from the air. Organisms in the biomass then remove both dissolved oxygen and organic materials from the film of wastewater. Further removal of dissolved oxygen and organic material occurs as the media continues rotation through the wastewater in the tank. Unused dissolved oxygen is mixed with the contents of the wastewater.

Using manufacturers standard performance curves it was estimated that a hydraulic loading of 3.33 gallons/day/ft.^{2} of media surface would produce the required effluent quality with a wastewater at 50°F. This design required the use of 4.8 million square feet of surface area. This would require the use of 48 shafts each containing 100,000 square feet of surface. The shaft dimensions are approximately 11'-10" diameter by 25 feet long.

A similar design was done on the other two secondary systems. The estimated construction costs, operation and maintenance costs and present worth values for the three secondary systems is summarized in Table 2. The costs include clarifiers and return sludge equipment where required.

SECONDARY TREATMENT COST ESTIMATES - 1976 DOLLARS
TABLE 2

Process	Construction Costs (Secondary Units Only)	Annual Energy Cost	Present Worth Plant (Liquid Handling)
Activated Sludge (Complete Mix Mode)	$3,024,300	$202,800	$12,981,250
Rotating Biological Contactor (Mechanical Drive)	$3,137,200	$ 97,300	$12,072,450
Activated Biofilter	$2,445,200	$156,600	$12,083,150


PRIMARY
SEDIMENTATION
SLUDGE
ROTATING BIOLOGICAL DISCS
SECONDARY
SEDIMENTATION
SLUDGE
TO
CHLORINATION

EAU CLAIRE URBAN AREA
FACILITIES PLAN

ROTATING BIOLOGICAL CONTACTOR

FIGURE 2

OWEN AYRES & ASSOCIATES

Based on the accuracy of the estimates all three treatment schemes proved to be cost competitive. To verify the assumptions made to arrive at the above estimates, pilot plants were run on both the RBC and ABF systems.

PILOT PLANT TESTING

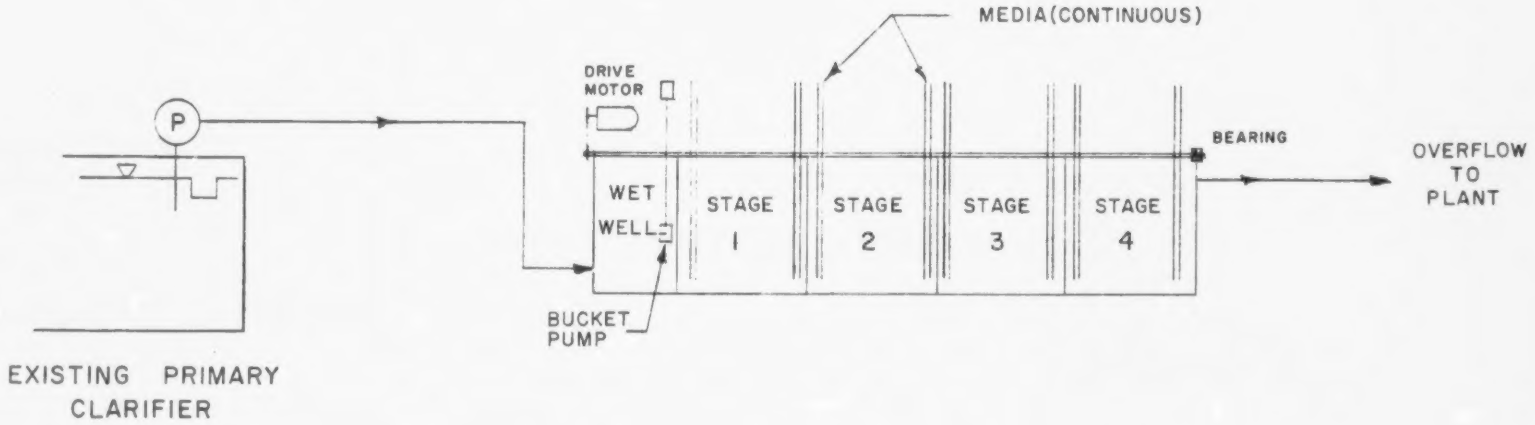
Two pilot plants were installed for testing. One plant consisted of RBC unit rented from the Autotrol Corporation, Milwaukee, Wisconsin. The second unit was fabricated at the City of Eau Claire wastewater treatment facility with equipment purchased from Neptune Microfloc, Inc. to establish a ABF system. Both units were constructed in the spring and summer of 1976.

The Autotrol pilot plant model rented is known as a two meter pilot plant. The overall dimensions of the unit were 14' x 8' x 8' and the active media surface area is 7900 sq. ft. Feed wastewater for this unit was obtained by pumping primary effluent from the City of Eau Claire municipal wastewater treatment plant to the unit wet well immediately preceding the media. From the wet well the wastewater was lifted into the unit by means of a bucket pump supplied with the pilot unit. The active surface area within the pilot plant was divided into four stages of equal surface area. The unit was driven by a motor connected with a gear reducer and chain system to rotate the discs. A schematic of the unit is shown by Figure 3.

The activated biological filtration system was constructed in an existing grit chamber that is not required for current plant operation. The tower dimensions were 4' square by 14' in height. Primary effluent was pumped into a wet well of dimensions 3' x 4' in depth. The water was then pumped into the top of the tower. The underflow from the tower returned into the wet well to be remixed with the incoming fluid.

Excess flow in the wet well overflowed into an aeration basin with dimensions 4' square by 7' deep. After aeration the wastewater flowed into a final clarifier of dimensions 8' long by 4' wide by 7' deep equipped with a sloped bottom and sludge hopper. The clarified liquid overflowed a V-notch weir and was returned to the municipal treatment plant flow. Sludge collected within this compartment was transferred back to the wet well by means of a submersible pump. A flow schematic is shown by Figure 4.

Both pilot plants were placed into operation and both produced an effluent quality much less than was predicted. To investigate whether the problem was one of process deficiency or wastewater composition, a bench scale, batch fed activated sludge unit was placed into operation. This unit, in two months of daily batch feeding on the twice per day basis, was unable to produce a substantial buildup of mixed liquor suspended solids. From the result it was concluded that the operational problems of both pilot plants were caused by a retardent present in the wastewater entering the Eau Claire treatment facility. A totally unexpected occurence based on a knowledge of the contributing area to the treatment plant.

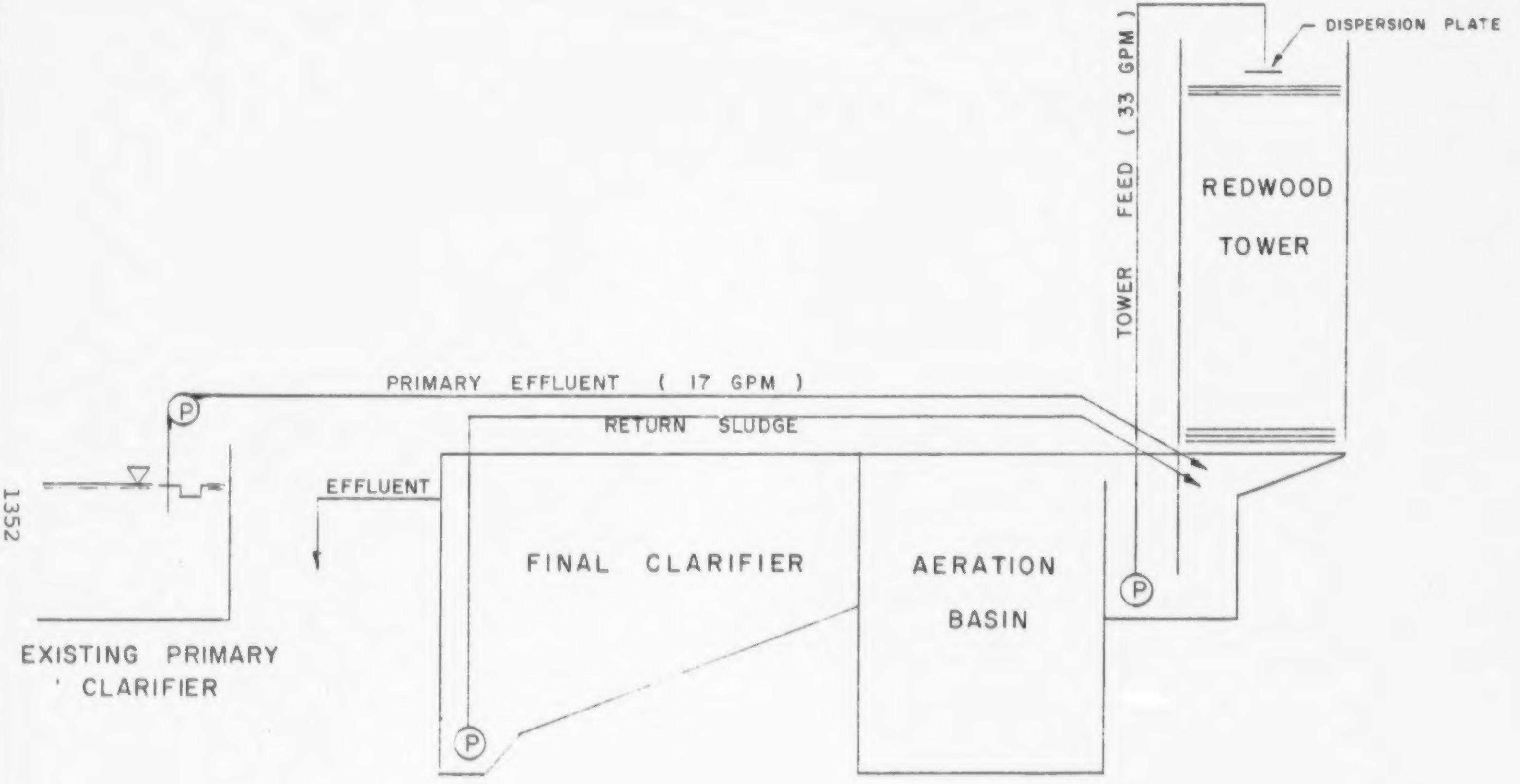

MEDIA(CONTINUOUS)
DRIVE MOTOR
P
BEARING
OVERFLOW TO PLANT
WET WELL
STAGE 1
STAGE 2
STAGE 3
STAGE 4
BUCKET PUMP
EXISTING PRIMARY CLARIFIER

ROTATING BIOLOGICAL CONTACTOR

PILOT PLANT SCHEMATIC

FIGURE 3

OWEN AYRES & ASSOCIATES


DISPERSION PLATE
TOWER FEED (33 GPM)
REDWOOD TOWER
PRIMARY EFFLUENT (17 GPM)
RETURN SLUDGE
P
EFFLUENT
FINAL CLARIFIER
AERATION BASIN
EXISTING PRIMARY CLARIFIER

ACTIVATED BIOLOGICAL FILTER

PILOT PLANT SCHEMATIC

FIGURE 4

OWEN AYRES & ASSOCIATES

In order to determine what wastewater constituent might be causing the lack of treatability, a sample of biomass from both the RBC and ABF pilot systems was collected for detailed laboratory analysis. The results of this testing indicated the potential for heavy metal toxicity. At this point it was decided to trace the metal contaminants to their source by sampling within the collection system. The City was divided into six smaller regions and sampling conducted. No unusual metal concentrations were found at any of the sampling locations.

However one location revealed the existance of extreme pH variations. These pH fluctuations were showing up in a lesser magnitude approximately four hours later in the influent to the wastewater treatment plant. The cause of the pH fluctuations was traced to a local industry which produces printed circuit boards. The industry was contacted and cooperated extremely well in controlling these severe pH fluctuations.

Both pilot plants were again activated in spring of 1977. The ABF unit still did not perform satisfactory. The RBC unit operated well at constant and varying flow rates. Analysis of the data collected showed the requirement for a design loading of 3.06 gpd/ft. This requires 5.4 million square feet of surface area or 54 large shafts. For symmetry reasons a design based on the use of 56 RBC shafts was recommended.

RBC DRIVE SYSTEM

The system design utilized a drive system consisting of two operating air compressors, piping and diffusers. The selection of the air drive system was based primarily on the maintenance requirements of 56 motors, gear boxes and variable speed drive boxes vs. the two air compressors and piping.

As an informational item, contractors bidding on the project were required to supply an alternate bid for a mechanically driven RBC system. Based on first cost a savings of $38,000 was possible by using the mechanical drive system. A present worth analysis was then conducted to determine the life cycle cost difference for the two systems. For reference purposes the mechanical drive system was used as a reference point (i.e. capital cost of mechanical drive = 0, capital cost of air drive = $38,000). The analysis is based on a total of 56 shafts with the following assumptions.

General Assumptions

- Electrical power costs. $0.0228/KWH (from 1976 Eau Claire Urban area Facilities Plan.) Cost in 1980 when the system is to be placed into operation is $0.00277/KWH.
- Labor Costs at $8.00/Hour.
- Interest rate 6-5/8%.
- Present Worth of annual sum (P/a) $\frac{.06625}{20} = 10.9099$
- Twenty year design life.

- Operating time is continuous.
- One horsepower - .746 KW.
- Shaft maintenance time (without drive) is 2.5 minutes/day/shaft.
- A 125 HP motor operating at 104 B.H.P. is 90% efficient.
- A 7-1/2 HP motor operating at 5 B.H.P. is 88% efficient.
- Blower maintenance time is 1/2 hour/day.
- Mechanical Drive maintenance time is 3 minutes/drive/day.

Air Drive System	vs.	Mechanical Drive System
Two blowers with a 125 HP motor rating are required.		Fifty-six drive motors of 7.5 H.P. are required.
Power draw is 104 B.H.P.		Power draw is 5 B.H.P.
A 90% efficiency can be expected at this loading.		An 88% efficiency can be expected at this loading.
Energy consumed is 86.2 KW per motor.		Energy consumed is 4.2 KW per motor.
Yearly power cost is $41,840.		Yearly power cost is $57,400.
Present worth of power is $456,000.		Present worth of power is $627,000.
Daily maintenance time is 3.3 hours.		Daily maintenance time is 5.1 hours.
Yearly maintenance cost is $9,640.		Yearly maintenance cost is $14,900.
Present worth of maintenance is $105,000.		Present worth of maintenance is $162,000.

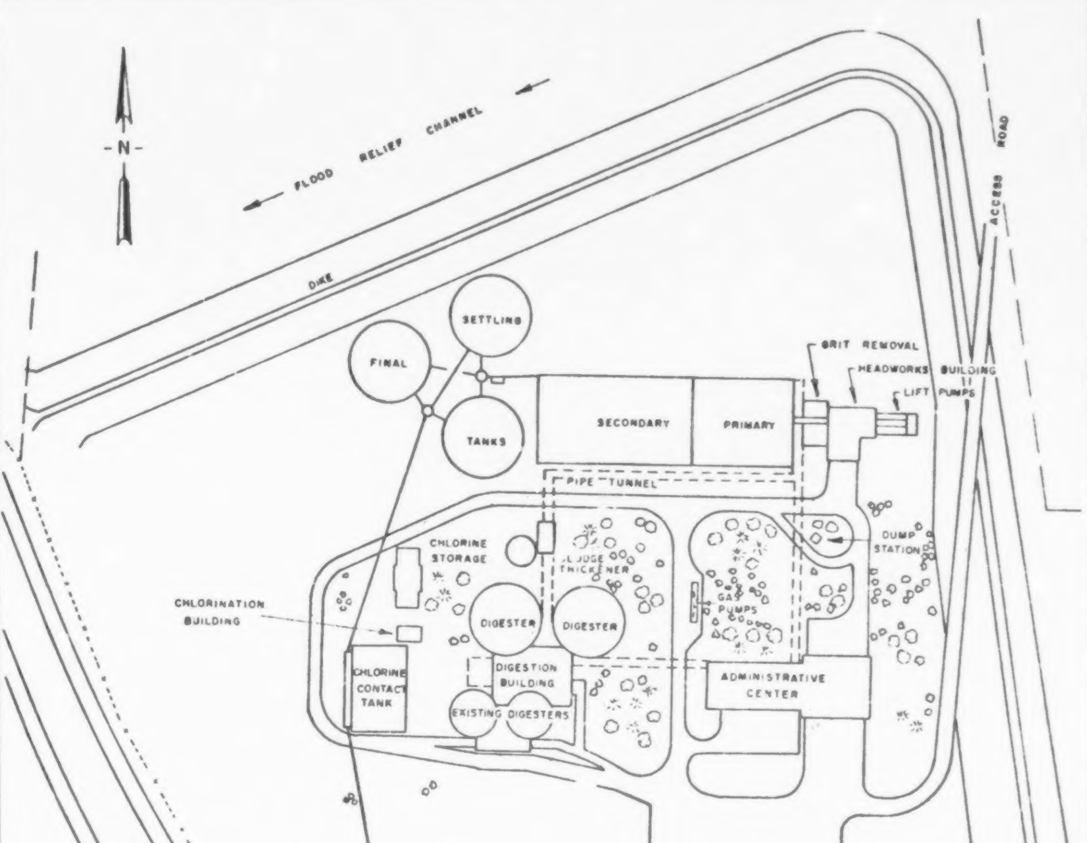
PRESENT WORTH COST SUMMARY
TABLE 3

Item	Mechanical Drive		Air Drive
Installation	0		38,000
Power	627,000		456,000
Maintenance	162,000		105,000
Total	$789,000		$599,000
Difference		$190,000	

Neither system requires any new technology based on operating components. Blowers, gear boxes, etc. have been used successfully in wastewater plants for years. Based on the cost analysis the air drive system was recommended for construction.

STATUS

The Eau Claire Wastewater Treatment Plant is currently under construction. An overall view of the plant is shown on Figure 5. The facility is expected to be placed into operation in the Fall of 1980. The secondary portion of the plant consists of 56 air driven rotating biological contactors followed by gravity sedimentation. Total estimated construction cost of the plant is $13,600,000.



N
FLOOD RELIEF CHANNEL
DIKE
SETTLING
FINAL
TANKS
SECONDARY
PRIMARY
GRIT REMOVAL
HEADWORKS BUILDING
LIFT PUMPS
ACCESS ROAD
PIPE TUNNEL
DUMP STATION
CHLORINE STORAGE
SLUDGE THICKENER
GAS PUMPS
CHLORINATION BUILDING
DIGESTER
DIGESTER
DIGESTION BUILDING
CHLORINE CONTACT TANK
EXISTING DIGESTERS
ADMINISTRATIVE CENTER
EFFLUENT LINE
FLOOD WATER PONDING AREA
I-94
I-94
DIKE
FLOOD PUMP LANDING
GATEWELL

EAU CLAIRE
WASTEWATER TREATMENT PLANT
SITE PLAN

FIGURE 5

OWEN AYRES & ASSOCIATES INC

AN EVALUATION OF THE COST-EFFECTIVENESS OF THE ROTATING BIOLOGICAL CONTACTOR PROCESS IN COMBINED CARBON OXIDATION AND NITRIFICATION APPLICATIONS

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INTRODUCTION

Environmental engineers are often confronted with effluent requirements calling for reductions in both biochemical oxygen demand (BOD_5) and ammonia nitrogen (NH_3-N) prior to stream discharge. A number of biological and physical-chemical processes are available for NH_3-N removal including ammonia stripping, breakpoint chlorination, ion exchange, suspended growth biological systems and attached growth biological systems. In municipal applications advanced BOD_5 removals (55%+) are virtually always achieved through the use of suspended growth or fixed growth biological reactors. For this reason, the most cost-effective process chains in applications calling for the simultaneous removal of BOD_5 and NH_3-N are typically completely biological.

This paper will compare the cost-effectiveness of the rotating biological contactor process to other biological processes for combined carbon oxidation and nitrification applications. Processes to be given consideration include: synthetic media trickling filters, conventional media trickling filters, and the single-stage activated sludge process. The evaluation will be based on a typical northern climate municipal wastewater.

BACKGROUND

All approaches to biological nitrification can be considered either *combined carbon oxidation-nitrification processes* or *separate stage nitrification* processes. They can be further subdivided into *suspended growth* or *attached growth* processes.

Suspended growth processes are those in which biological solids are suspended in a liquor by some mixing mechanism. The conventional activated sludge process is the most common example of a suspended growth process. In attached growth processes, the bulk of the biomass is fixed to some permanent media held in the reactor. The trickling filter and the Rotating Biological Contactor are common examples of attached growth processes. The suspended growth processes require post-clarification and the return of the biological solids to the head of the reactor to maintain an adequate biomass inventory. Attached growth systems will also generally require post-clarification for compliance with effluent requirements; however, since an adequate inventory of biomass remains fixed to the reactor's media recycle for the purpose of inventory maintenance is not required. Partial recycle of clarified effluent is practiced in some attached growth processes, most notably in trickling filtration, but to meet other requirements or achieve other technical advantages.

In a separate stage nitrification process, the BOD_5 and NH_3-N removal functions are largely confined to separate reactors. The carbon removal stage precedes the nitrification stage to lower the organic load and to facilitate the development of a population of nitrifiers in the second stage. Intermediate clarification is essential when a suspended growth first stage is employed. It is optional when an attached growth first stage is employed.

DESIGN CONSIDERATIONS AND CRITERIA

A cost-effectiveness analysis of alternative nitrification processes cannot be undertaken without first discussing design criteria. The following paragraphs review design considerations for both suspended growth and attached growth processes.

Conventional Media Trickling Filtration

The ability of conventional media trickling filters to provide nitrification in a single-stage is not widely recognized in this country. The authors have had conversations with individuals in the regulatory, research and engineering communities who have argued that

A) Deep beds are needed (20 feet +) to provide complete nitrification,

B) Nitrification cannot be effected in conventional trickling filters because of low wastewater residence time, and

C) Nitrification can only be effected if the influent to the trickling filter is less than 30 to 40 mg/liter

An extensive review of the literature belies the above assertions and demonstrates the efficacy of conventional trickling filtration as a combined carbon oxidation-nitrification process. Further, it is possible to develop empirical design criteria based upon available pilot- and full-scale operating data.

Between 1944 and 1961 a number of articles were written summarizing investigations into nitrification in conventional media trickling filters. To be of utility to the current Paper, a researcher must have reported at least the following: observed influent and effluent NH_3-N concentration, influent BOD_5 concentration and hydraulic load, and the type/volume of media in the filter. Further, only investigations that loaded filters with municipal primary effluent were considered appropriate for use. Four significant research efforts were found that provide required information: 1) the National Research Council's investigation[1] of twelve operating full-scale, single-stage filters across the United States in 1944; 2) Stones' extensive 1961 investigation[2] of large-scale pilot plants at Salford, England; 3) Grantham's pilot-scale studies[3] at Gainsville, Florida in 1953; and 4) Burgess' pilot-scale work[4] at Corvallis, Oregon in 1961. The body of data collected during these research efforts covers a wide range of organic loadings, climatic conditions, operating modes and types of filter media.

The empirical design criteria proposed herein will relate percent NH_3-N removal to the pounds of BOD_5 applied per day per 1000 ft^2 of filter media. Such a criteria is not only convenient but it also has a sound conceptual basis. The amount of biomass held in a reactor should generally be portional to the amount of surface area provided by the media in the reactor. A precise estimate of the biomass available within an attached growth reactor would require knowledge of the thickness of the slime coating the media, the effective media surface area and the fraction of volatile versus inert solids in the slime. As detailed information of this nature is unavailable, a direct proportionality between biomass and media surface area must be presumed. At high influent BOD_5 loads, heterotrophic organisms can be expected to dominate the slime found on the media. As the ratio of BOD_5 load to the available surface area decreases, nitrifying organisms can begin to dominate the slime beginning with the lowest layers of the media. Several hypotheses can be advanced to explain the reduced inventory of nitrifying organisms at the higher BOD_5 loading rates and in the upper layers of filters under moderate loading rates. One attractive hypothesis is based on the observed natural variation in growth rates of the heterotrophic and nitrifying organisms responsible for BOD_5 and NH_3-N removal. The heterotrophic organisms grow more rapidly and have a greater yield than the nitrifiers. The higher sluffage rate occurring during BOD_5 removal causes a continual displacement of the nitrifiers from the film preventing the establishment of an adequate nitrifying population. This scenario is analygous to an activated sludge process operated at comparatively low mean cell residence times where the nitrifying organisms are being wasted at a faster rate than they can regenerate. A second reasonable hypothesis is based on the relationship of the availability of dissolved oxygen to the rate of ammonia oxidation. The rate of ammonia oxidation decreases with decreases in dissolved oxygen concentration. The nitrifiers are more sensitive to depressed dissolved oxygen concentrations than the heterotrophic organisms responsible for carbon oxidation. In filters with higher BOD_5 loadings, lower dissolved oxygen levels might logically be anticipated and thus the nitrifiers are again put at a competitive disadvantage. Both of the above hypotheses are speculative and in any event a firm definition of the operable mechanism or mechanisms is not critical to this Paper as an empirical design criteria is being proposed.

The authors have re-expressed the work of the National Research Council and Stones on a pounds BOD_5 applied per 1000 ft^2 of media basis and have plotted the data on Figures I and II. The work of Stones implies that recirculation has a beneficial effect on nitrification. A 1:1 recirculation ratio was employed. The National Research Council's data confirms Stones' observations in the lower loading ranges and provides information on the effect of higher organic loads on nitrification. In general, each of the National Research Council data points represents the average of four months of data at an individual plant. Plants from both cold and warm weather climates and plants with and without recirculation were covered in their survey.

Figures I and II, along with information from Gainsville and Corvallis, was combined with more recent unpublished information that the authors' have in their files from five additional full-scale rock trickling filters to prepare Figure III titled "Basis of Design: Conventional Media Trickling Filters". It is proposed that the upper boundary of the envelope of performance be employed in applications involving warm wastewaters and involving recirculation. The lower boundary is to be employed in applications involving cold wastewaters and contemplating no recirculation. In all cases, single-stage filters with minimum bed depths of six feet may be presumed.

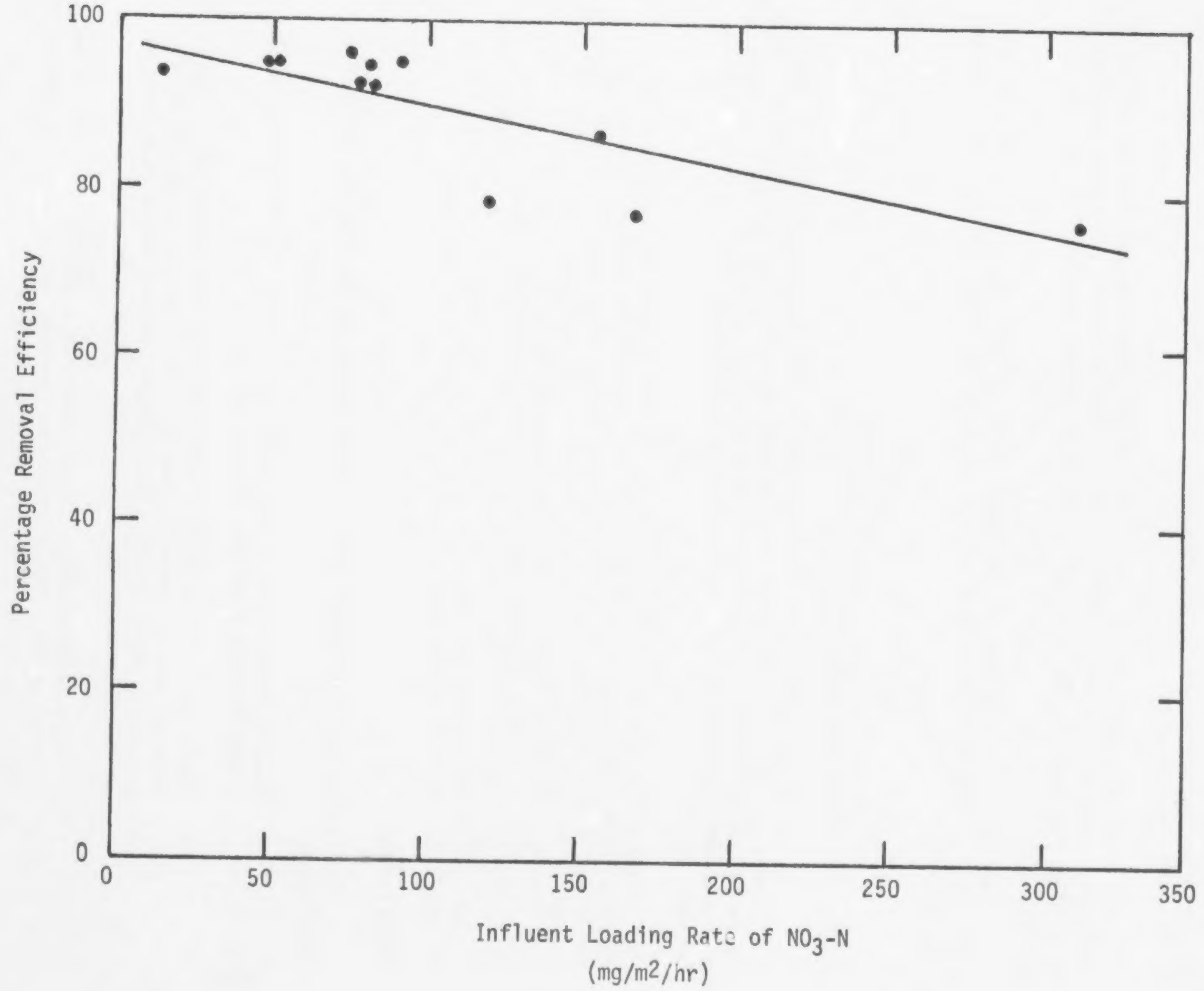
Figure III shows that loadings lower than 0.5 $\#BOD_5/1000\ ft^2 \cdot day$ are necessary to achieve 90% NH_3-N removal.

Synthetic Media Trickling Filters

Synthetic plastic media was originally developed for trickling filters in the mid 1950's. Because of its relatively high ratio of surface area to volume and because of its increased void space, it gained acceptance in the early 1960's as a recognized substitute for conventional media. There are two prominent types of synthetic media in common use: 1 - corrugated PVC sheet modules which are cut on-site and then laid into the filter bed, and 2 - plastic rings which are dumped into the filter bed.

In attempting to establish a design criteria for combined carbon oxidation-nitrification in synthetic media trickling filters the authors found virtually no published research or full-scale operating data. Most work to date has assumed that the influent feed to the nitrification reactor would be secondary effluent or its equivalent. An attempt is made in the following paragraphs to develop a design criteria for combined carbon oxidation-nitrification synthetic media reactors through the use of available information.

FIGURE 15: Two Stage Removal Efficiency



nitrogen concentrations of 22 mg NO_3-N/1.

The data shown in Figure 11 indicates that the nitrate removal efficiencies in the RBC unit was a function of both nitrate loading rate, and the carbon to nitrate ratio used. In this study, for the loading rates investigated, the optimal methanol to nitrate ratio varied between 2.4 - 3.0, depending on nitrate loading rate.

Although an exact solids production monitoring balance was not continually conducted, the solids production was low. This is to be generally expected in anaerobic processes. The average value of effluent suspended solids was 3.4 mg/1.

Temperature dependency in the range of 6 to 16°C was also evaluated in this study. The results indicate that it could be described by the following expression:

$$K_T = K_{20}\ [1.075^{(T-20)}]$$

B - Two Stage Pilot Unit

The two stage pilot plant was operated at the Marlborough W.T.P. during the summer and fall of 1977, one year after the single stage operation. Each stage was essentially the same in surface area and volume as the one stage unit. his unit was operated at a higher rotational speed and had a greater spacing between discs.

Figures 12, 13, 14, and 15 show the curves of nitrate loading rates vs. removal rates and nitrate loading rates vs. percentage removal rates for the two stage unit. Figures 12 and 13 present first stage results while Figures 14 and 15 present the two stage unit results. To compare the effects of staging for the two stage investigation in 1977, Figures 13 and 15 may be examined simultaneously. These figures indicate that at an application rate of 1 lb. of NO_3-N per 1000 ft^2 of surface area,(205 mg NO_3-N/m^2-hr), the single stage would remove 63 percent of the nitrate while a two stage unit would remove 83 percent. In both instances, the contact times were the same.

CONCLUSIONS

The following conclusions were reached based on the studies described in this paper:

1. Denitrification of a nitrified wastewater can be readily accomplished using rotating biological contactors.
2. There is a linear relationship between the amount of nitrate applied and the amount removed. This can be expressed by the equation; NR = .59 NA + 18.8, with the variables expressed as mg NO_3-N/sq m-hr.
3. Nitrate concentration in the unit has a significant effect on removal rates. The higher the level of nitrate within the unit, the higher the removal rate will be.

4. 90% nitrate reduction can be expected at loading rates of up to 55 mg NO_3-N/sq m-hr under normal operating conditions in a single stage reactor. For a two stage reactor, 90% nitrate reduction will occur at loading rates up to 100 mg NO_3-N/sq m-hr.

5. Removals will increase with increasing detention times for any unit at a given influent nitrate concentration.

6. Methanol requirements for optimum removal can be estimated as: CH_3OH mg/l = 2.6 NO_3-N mg/l + 0.91 D.O. mg/l. Lesser amounts limit removals and excess amounts produce high effluent BOD values.

7. The temperature dependency can be estimated as:

$$K_T = K_{20}\ 1.075^{(T-20)}$$

8. Solids production was low at all loading rates during these investigations.

9. Staging will improve the performance of RBC denitrification units.

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NITROGEN REMOVAL IN ROTATING BIOLOGICAL CONTACTORS WITHOUT THE USE OF EXTERNAL CARBON SOURCE

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Introduction

The main pollution problem resulting from wastewater discharges in Norway is eutrophication of the inland lakes and the fjords. In order to fight eutrophication, phosphate removal is now carried out in most treatment plants. The role of nitrogen in eutrophication has been more or less neglected, and treatment plants for nitrogen removal are non-existant in Norway. The interest in nitrogen removal is however growing caused by the fact that nitrogen has been found to be growth limiting nutrient in some eutrophic lakes.

The interest for nitrogen removal combined with the rising popularity of the biodisc in Norway, has initiated research at The University of Trondheim on the use of biodisc-plants for the removal of both organic matter, phosphorus and nitrogen.

NITROGEN REMOVAL IN BIODISC-SYSTEMS

The literature on biological nitrogen removal in general is vast, at least compared to the number of actual plants that have been built throughout the world. The majority of the papers on nitrification/denitrification, however, deals with suspended cultures. Most of the denitrification studies that have been carried out with fixed film cultures, have dealt with anaerobic filters or lately, fluidized beds. In the last few years some authors have reported from investigations of denitrification in biodisc systems /1, 2, 3, 4/.

The flow diagram normally used when biodiscs are to be used for both nitrification and denitrification, is the one shown in Fig. 1.

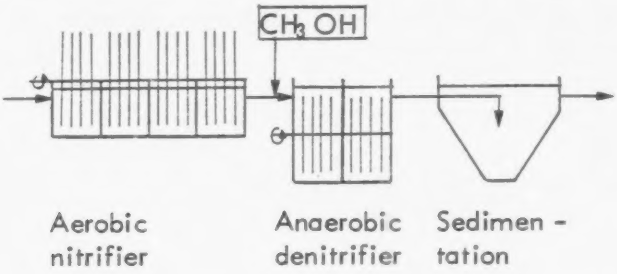


Fig. 1. Normal process for nitrogen removal in biodisc plants.

The advantage of this system compared to a suspended culture system or to a system with anaerobic filter as denitrifier, is that only one sedimentation unit is needed.

The main disadvantage of the system, is the need for an additional carbon source to obtain denitrification. As in other denitrification systems, methanol would normally be used. There are several disadvantages with methanol addition:

1. It increases operation costs. The enormous increase in methanol price the last few years, has made this point very important.
2. It increases investment cost since an advanced control system is necessary in order to add the correct amount of methanol.
3. Surplus methanol not used for denitrification would increase the effluent BOD. The amounts normally added corresponds to 80 - 100 mg BOD/l in municipal sewage, which could be detrimental to the receiving water if denitrification for some reason did not take place or in case of failure in the control system mentioned above.

The goal with our research was to unnecessitate the need for an external carbon source and use the raw water itself as such. This has successfully been done in suspended culture systems in the so-called Bardenpho-process proposed by Barnard /5/, where nitrate-rich mixed liquor from an activated sludge tank, is recycled to an anaerobic suspended culture tank ahead of the aerobic one.

In the process presented in this paper, the same principle has been utilized in a biodisc-system, see Fig. 2.

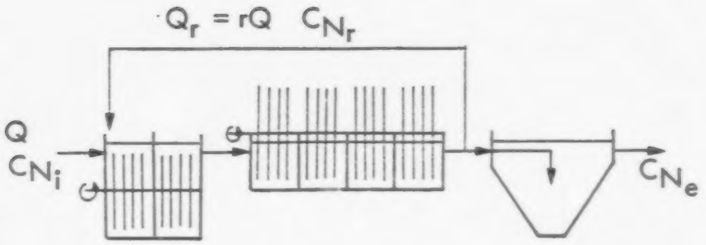


Fig. 2. Proposed process for nitrogen removal in biodisc plants.

The ammonium-rich influent passes through a tank with submerged biodiscs where anoxic conditions will exist. From this tank it enters an ordinary aerobic biodisc unit where nitrification takes place. Nitrified water is then recycled to the inlet of the anoxic tank where denitrification takes place and where raw water can act as carbon source.

The maximum possible nitrogen removal that can be achieved with the process, may be calculated from a mass balance assuming complete nitrification and denitrification in the two steps:

$$Q\,C_{N_i} = rQC_{N_r} + QC_{N_e}$$
$$= rQC_{N_e} + QC_{N_e} = QC_{N_e}(1+r)$$

Removal:

$$\frac{C_{N_i} - C_{N_e}}{C_{N_i}} \cdot 100\% = \frac{C_{N_i} - \frac{C_{N_i}}{r+1}}{C_{N_i}} \cdot 100\% = \frac{r}{r+1} \cdot 100\%$$

where C_{N_i}, C_{N_e} and C_{N_r} are nitrogen concentrations in the influent, in the effluent and in the recycled flow respectively.

A removal of 90% requires for instance a recycled flow of 9 times the influent flow,while 80% is achievable with a recycle ratio of $r = 4$.

EXPERIMENTAL ARRANGEMENT

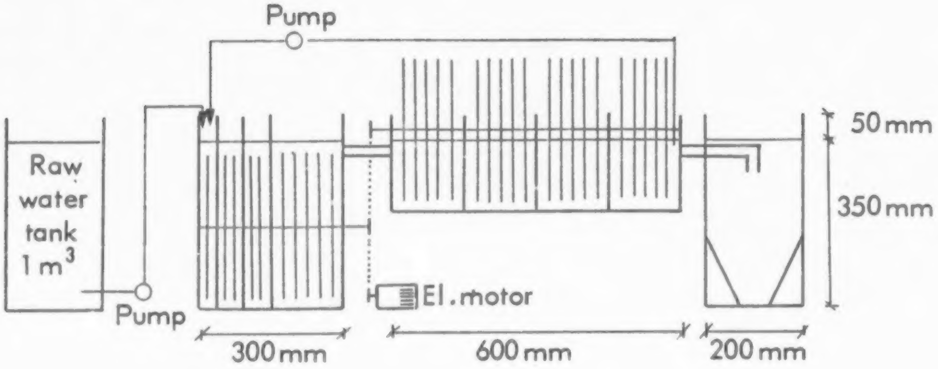
The experiments were carried out in two plexi-glass biodisc units arranged in parallel, each with a capacity of 1 - 10 l/h (Fig. 3).

Experiments with three different types of raw water have been carried out:

1. Artificial wastewater
2. Municipal wastewater (sewage)
3. Leachate from sanitary landfill

Artificial wastewater

The reason for starting out with artificial wastewater was to ensure influent characteristics to be as stable as possible in order to investigate the possibilities of the process. The water was composed to be as similar to municipal wastewater as possible.



	Denitrifier	Nitrifier
Disc diam.	300 mm	300 mm
Disc number	10	20
Comp. number	4	4
Disc area	1.41 m^2	2.82 m^2
Tank volume	32.8 ℓ	23.6 ℓ
Rotational speed	15-16 rev/min	15-16 rev/min

Fig. 3. Experimental unit.

Table 1 Composition of artificial waste water

Glucose	: 200 mg/l	$KHCO_3$	: 50 mg/l
Yeast extract	: 30 "	$NaHCO_3$	: 130 "
Dried milk	: 120 "	$MgSO_4 \cdot 7\,H_2O$	: 50 "
Urea	: 30 "	$FeSO_4 \cdot 7\,H_2O$	: 2 "
NH_4Cl	: 60 "	$MnSO_4 \cdot H_2O$	: 2 "
$Na_2PO_4 \cdot 12\,H_2O$	: 100 "	$CaCl_2 \cdot 6\,H_2O$	: 3 "
		Bentonite	: 40 "

This water gives the following results when analyzed:

Total COD (COD_t) : 400 mg O/l
Soluble COD (COD_s) : 300 "
Soluble BOD (BOD_{5_s}) : 188 "
Tot P : 9 mg P/l
Tot N : 36 mg N/l
Suspended solids : 90 mg/l
Alkalinity : 130 mg $CaCO_3$/l

Sewage

The sewage was taken from an apartment house during morning hours in order to have high ammonium concentrations. The water showed, however, after presettling, to have a relatively low soluble organic content ($COD_{filtered}$ ≈ 100-160 mg O/l) compared to the ammonium content (NH_4-N ≈ 35 - 65 mg N/l). As will be commented later, this gave us some problems since COD became the limiting factor for denitrification. In part of the experiments this relationship between COD and N was corrected for by introducing dairy effluent (dried milk) to the sewage.

Leachate from Landfill

The high amount of ammonium in leachate may cause both oxygen deficit and eutrophication in the receiving waters, and a couple of such situations in Norway have intensified research on treatment of leachate from landfills.

A review through the literature /6/ will show that treatment of leachate is very troublesome, and that anaerobic treatment has been used with some success. We found it interesting, therefore, to test the proposed process on leachate.

The leachate used in this experiment was that of the landfill of Trondheim. The experiments were carried out during a period of heavy rainfall that diluted the leachate. Table 2 shows composition of the leachate during 1975 and 1976 compared with the composition during our experiment.

Table 2 Composition of leachate

Parameter	1975 - 76	Actual experiment
COD mg/l	1250 - 1650	400 - 800
Tot N	80 - 100	80 - 130
NH_4-N	65 - 80	70 - 125
Tot P	0.7 - 0.8	0.6
Fe	105 - 140	100 - 145
pH	6.8	7.0

The difference is especially significant for COD while the difference for the other parameters are minor.

In order to prevent the problems with ironhydroxide-precipitation on the biofilm, most of the iron was precipitated in advance by adjusting the pH to 8, aeration and filtration.

RESULTS

Nitrogen removal

In Fig. 4 are shown experimental results from situations where nitrification and denitrification limited by any factor. The results are compared with the theoretically possible removal as mentioned earlier.

The figure shows that the results fit the theoretical removal quite nicely. This demonstrates that given the situation that no factor (such as pH, O_2-concentration, BOD-loading, toxic substances, and so on) limits nitrification or denitrification, the proposed process will work as anticipated. Concequently, the process is an interesting alternative to the different nitrogen removal processes used today.

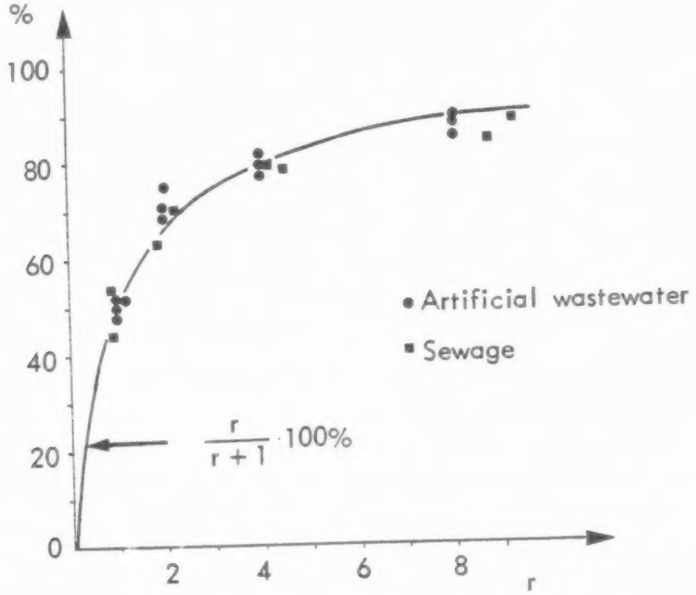


Fig. 4. Nitrogen removal against resirculation ratio.

Nitrification

Several factors are influencing the degree of nitrification, such as content of biodegradable organic matter, oxygen, pH and temperature.

Most workers agree that the oxygen content in the water may be a limiting factor with regard to the nitrification rate. Weng and Molof /7/ found that nitrification took place only if more than 2 mg O_2/l was present. In our experiments the oxygen content was normally in the range of 3-6 mg O_2/l and never below 2.5 mg O_2/l in the nitrifier.

Optimum pH for nitrification has been found to be around 8 /8/, but nitrification is achieved in the range of pH = 6.5 - 9.5. In all our experiments the pH was in the range of 7.0 - 8.3.

Optimum temperature for nitrification has been found to be in the range of 28 - 36 °C /9/ where municipal treatment processes seldom operate. Antonie /2/ has postulated that the nitrification rate is constant over 15 °C, but decreases with decreasing temperature under that. Pretorius /10/ found Q_{10} $(= \hat{\mu}_{20}/\hat{\mu}_{10}) = 1.25$ in his biodisc experiments.

No temperature control was used in the experiments presented here. During the artificial waste water period, the temperature was in the range of 13 - 16 °C, while it varied more in the sewage and the leachate period (10 - 16 °C).

In order to achieve nitrification, the content of biodegradable matter has to be brought down sufficiently.

Prior to the experiments with the proposed treatment process (Fig.2), some experiments with the process shown in Fig. 1 were performed in order to investigate the acceptable organic loading to achieve nitrification. These experiments were carried out with the artificial wastewater.

Fig. 5 shows the relationship between the degree of nitrification (per cent removal of NH_4-N) and the content of soluble organic matter ($COD_{filtered}$) in the effluent.

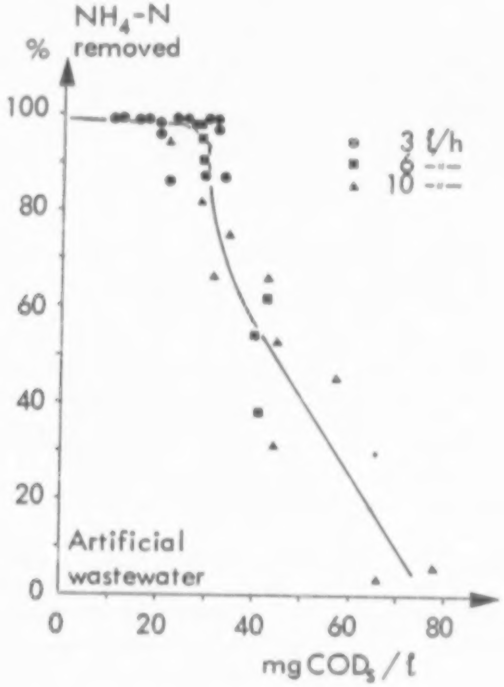


Fig. 5. Effect of soluble COD on the degree of nitrification.

It is demonstrated that a soluble COD-content of less than 30 mg COD/l was needed to ensure complete nitrification. Antonie /2/ reported that nitrification started when the BOD-concentration was brought down to 30 mg BOD_5/l, while it was fully developed below 8 - 10 mg BOD_5/l. Weng and Molof /7/ found that nitrification occurred at COD-concentrations below 50 mg COD/l. Both of these findings correspond to a soluble COD-concentration of about 25 - 35 mg COD/l and are in agreement with our results.

In Fig. 6 is plotted the degree of nitrification against organic loading of soluble COD per unit area of biofilm. It is demonstrated that the load had to be kept below 25 g $COD_s/m^2 \cdot d$ to ensure full nitrification.

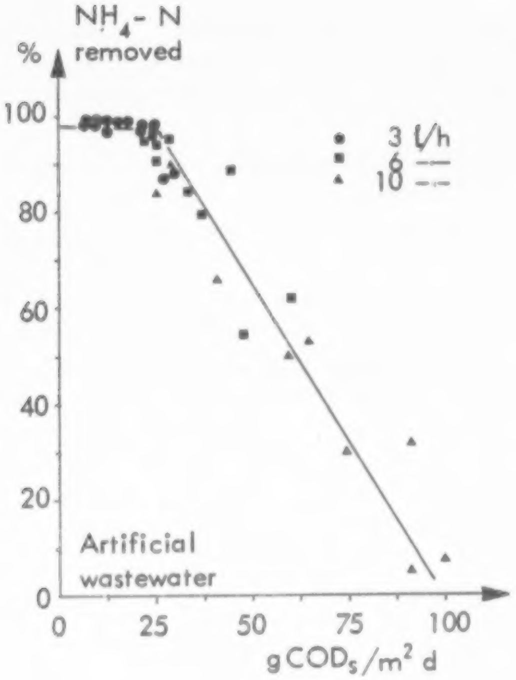


Fig. 6. Effect of organic loading on the degree of nitrification.

To evaluate the results on nitrification in the experiments with the proposed anoxic-aerobic process is a little complicated because the recirculated water contributes both to the COD-loading and the ammonium-loading. The COD that enters the nitrifier is probably composed of one easy biodegradable fraction from the raw water and one heavy biodegradable fraction from the recirculated water.

In the sewage experiment the BOD_5 in the recirculated water for one run was found to be as low as 3.5 mg BOD_5/l, while the COD was 39 mg COD/l. It may thus be stated that the organic load from the recirculated water does not contribute much to the total biodegradable organic load that will determine the degree of nitrification.

In the experiments presented in Figs. 5 and 6, recirculation was not used. In the experiments with the proposed process higher COD_S-values could be tolerated to ensure nitrification. This is partly due to the fact that non-biodegradable COD is recirculated. In the leachate experiments, nitrification took place even at COD-values of 150-200 mg COD/l.

In Fig. 7 is plotted the nitrification rate against ammonium-loading for situations where nitrification took place (was not limited by organic loading). The spread in the results may be explained by the lack of temperature control.

It is interesting to note that for all the types of raw water tested, there seems to be a critical loading of about 100 - 150 mg NH_4-N/m^2·h over which the nitrification rate is at its maximum and independent upon the loading. This maximum nitrification rate is about 100 - 120 mg NH_4-$N_{removed}$/m^2·h both for the artificial wastewater and the sewage, while it is considerably lower for the leachate, around 40 mg NH_4-$N_{removed}$ per $m^2 \cdot h$.

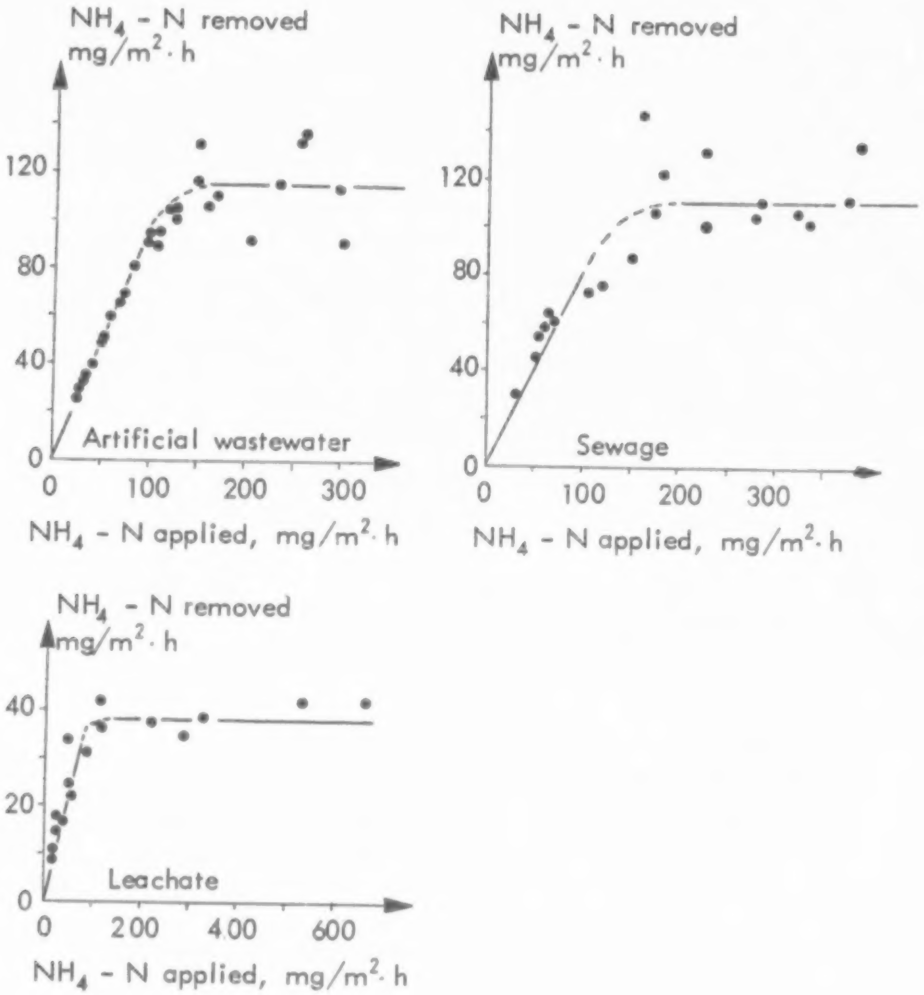


Fig. 7. Nitrification rate against ammonium-loading.

In Fig. 8 is shown the nitrification rate in each step of the unit, which may be assumed to be complete mixed.

Even though the spread in the results are great, it may be concluded that the nitrification process follows Monod-kinetics, with zero-order kinetics at higher ammonium concentrations and first order at lower concentrations. It seems that zero order kinetics governs the process at concentrations higher than 4-6 mg NH_4-N/l in normal sewages and higher than 20 mg NH_4-N/l in landfill leachate.

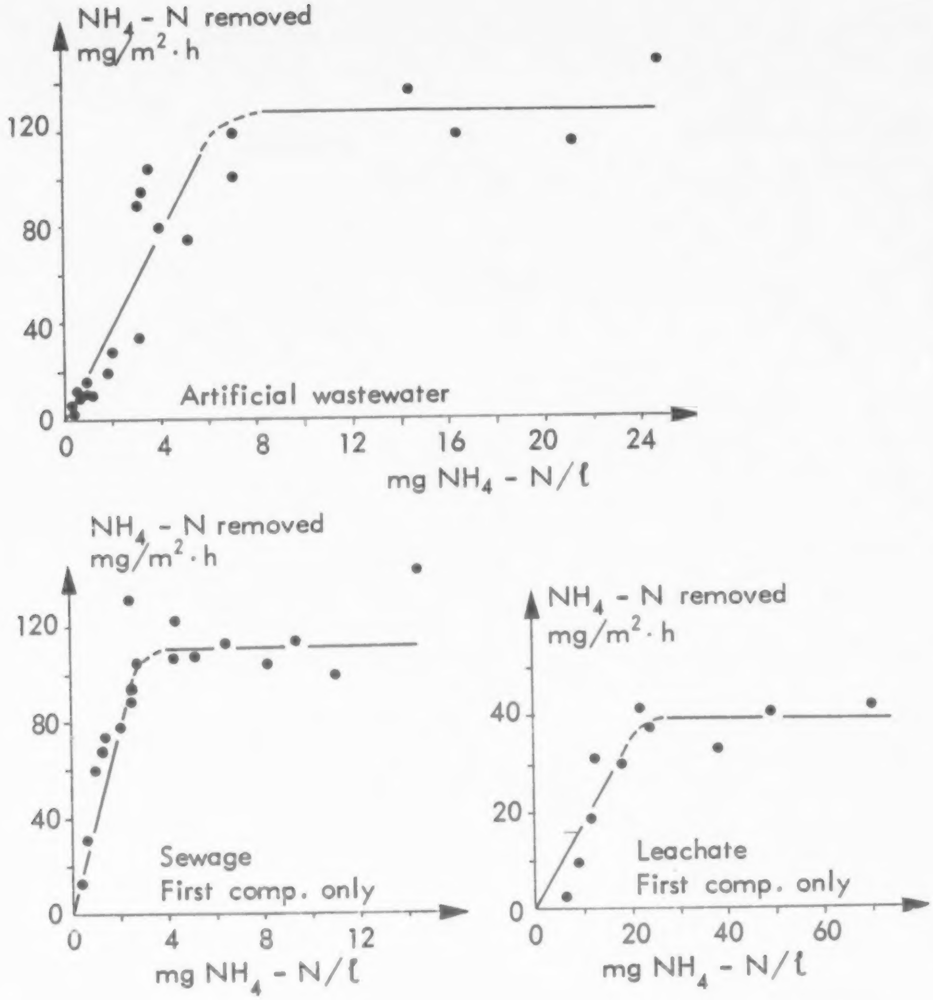


Fig. 8. Nitrification rate against ammonium-concentration.

In Table 3 the nitrification rates found in this study is compared to what has been found in other studies.

Table 3 Nitrification rates

Raw water	Nitrification rate	Reference
Sewage with $NH_4(HCO_3)$ addition	77.5 mg NH_4-N/m²·h	Pretorius /10/
Municipal sewage	44 mg TKN/m²·h	Murphy et al./11/
Artificial wastewater	max. 120 mg NH_4-N/m²·h	This study
Municipal sewage	max. 110 mg NH_4-N/m²·h	" "
Landfill leachate	max. 40 mg NH_4-N/m²·h	" "

Denitrification

Several factors are influencing denitrification, such as oxygen concentration, pH, temperature, carbon source, nitrate-loading and ratio between loading of organic matter and nitrate.

The denitrification has to be carried out at anoxic conditions. True anaerobic conditions in the liquid does not seem to be necessary /12,4/. In the experiments presented here, conditions were normally anaerobic but denitrification took place even at oxygen-concentrations in the water around 1 mg O_2/l. It was not necessary to cover the free water surface to the atmosphere.

Optimum pH for denitrification seems to be pH = 7 - 7.5 /8/. In our experiments the pH varies between 7 and 8 in the wastewater experiments and between 8 and 8.5 in the leachate experiments.

Methanol is normally used as carbon source. Several authors have investigated nitrate/nitrite-removal as a function of methanol consumption for fixed film systems. Christensen and Harremoes /12/ reviewed the different investigations, and concluded that empirically, a methanol to nitrogen ratio of 3 g CH_3OH/g NO_3-N was necessary to reach more than 90% nitrate removal.

In suspended cultures the theoretical methanol consumption would be 2.47 g CH_3OH/g NO_3-N /13/, which is close to what has been measured by many authors /12/.

The most interesting alternative to methanol is undoubtedly the raw sewage itself. Narkis, Rebhun and Sheindorf /14/ have recently published results from suspended culture experiments where different carbon sources (methanol, sodium acetate and chemically treated raw sewage) were used. They concluded that by expressing the organic matter as soluble BOD, a critical value of 2.3 mg BOD/mg NO_x-N existed when 100% denitrification was to be reached regardless of what carbon source was used.

In the experiments presented here, soluble COD was, for different reasons, used as a measure for the organic matter. In the artificial wastewater experiments the COD-concentration did not seem to be limiting for denitrification. In the municipal wastewater experiments, sewage from morning hours was used. This had an abnormal high nitrogen concentration compared to the soluble COD-concentration. Complete denitrification was consequently not obtained. In Fig. 9 the nitrate removal is plotted against the COD/NO_x-N ratio (COD in raw water and total NO_x-N applied).

It is demonstrated that almost complete denitrification was obtained when the ratio of soluble COD in the raw water to the NO_x-N concentration into the anaerobic unit was over 7, and that the relationship between the removal and this ratio was linear for ratios lower than 7.

This made us add dried milk corresponding to a soluble COD addition of 100 mg COD_S/l in the proceeding experiments. (Bio-disc systems are, by the way, normally used in Norway when dairy effluents are treated together with municipal sewage.)

The fact that the influent COD is reduced through the anoxic step is another valuable aspect about the proposed process. Reduced organic loading on the aerobic step makes it possible to reduce the necessary area in this step in order to achieve the required COD to obtain nitrification. Altogether, this means lower investment cost compared to a situation were the denitrifier is proceeding the nitrifier.

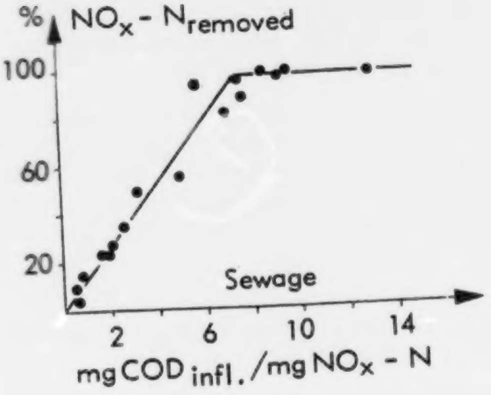


Fig. 9. Degree of denitrification against COD/NO_x-N ratio.

In order to find how much COD that was consumed by denitrification, the ratio between $COD_{consumed}$ and $(NO_x\text{-}N)_{consumed}$ was calculated. This ratio should be expected to be constant for the different runs. In order to visualize the results the ratio is plotted against recirculation ratio. Even if there is a considerable spread in the results, Fig. 10 shows that 2.5 - 3.5 mg COD is consumed for each mg NO_x-N removed by denitrification.

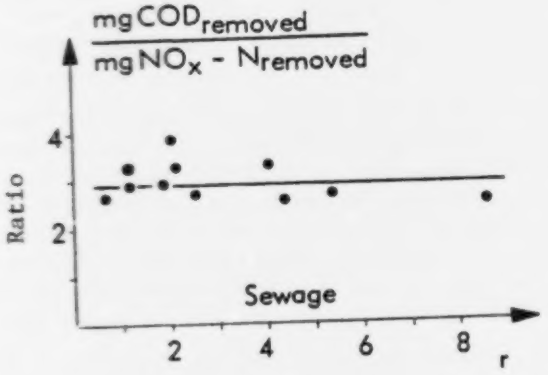
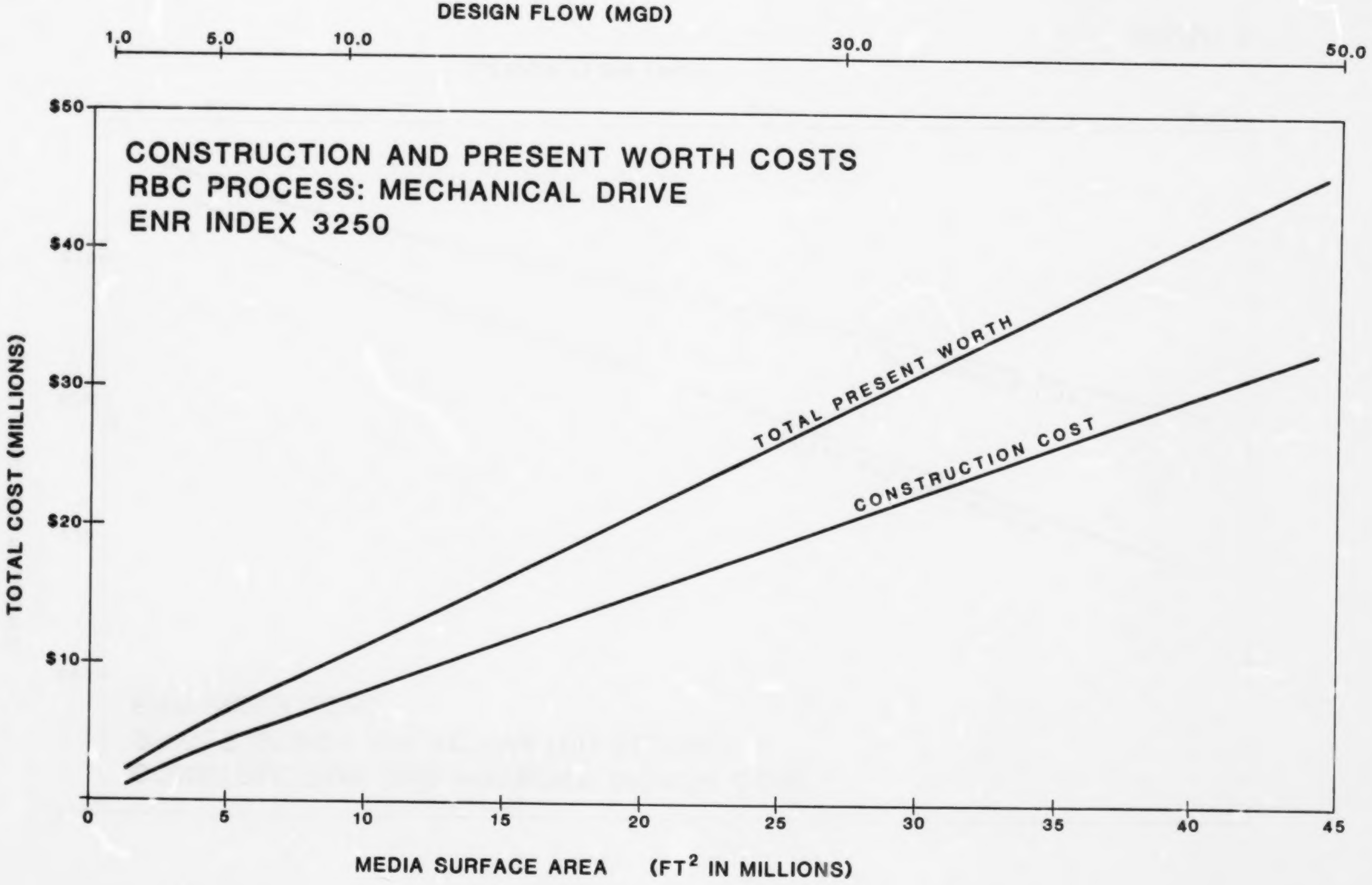


Fig. 10. COD consumed during denitrification.

In Fig. 11 are plotted the NO_x-N removal (%) versus NO_x-N loading for the different waters.

In the artificial wastewater experiment, a maximum load of 4 g NO_x-N/m²·d could be tolerated in order to achieve more than 90% denitrification. The corresponding load in the sewage experiment seemed to be a little bit higher, 6 g NO_x-N/m²·d.



CONSTRUCTION AND PRESENT WORTH COSTS
RBC PROCESS: MECHANICAL DRIVE
ENR INDEX 3250

FIGURE IX

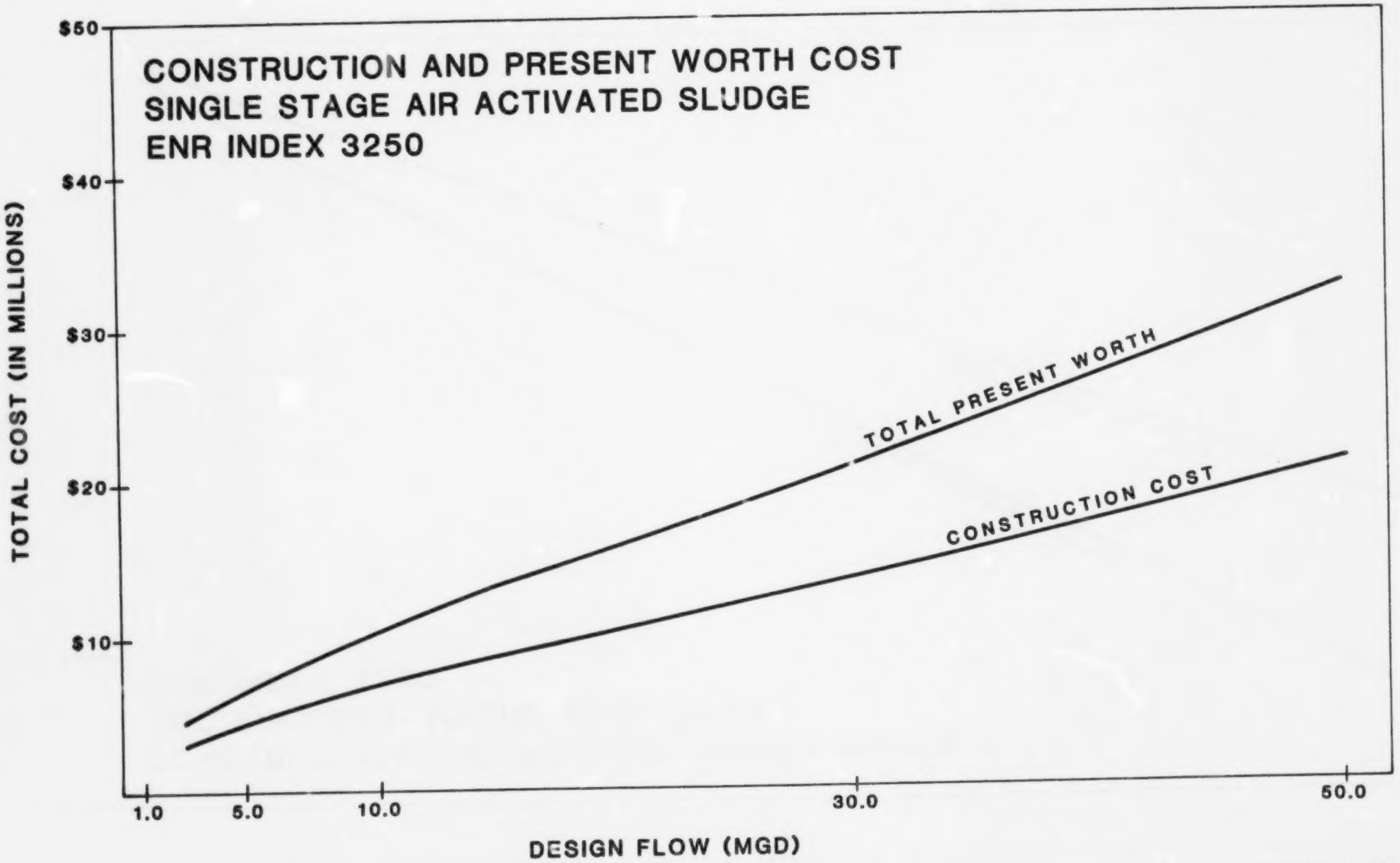

CONSTRUCTION AND PRESENT WORTH COST
SINGLE STAGE AIR ACTIVATED SLUDGE
ENR INDEX 3250

FIGURE X

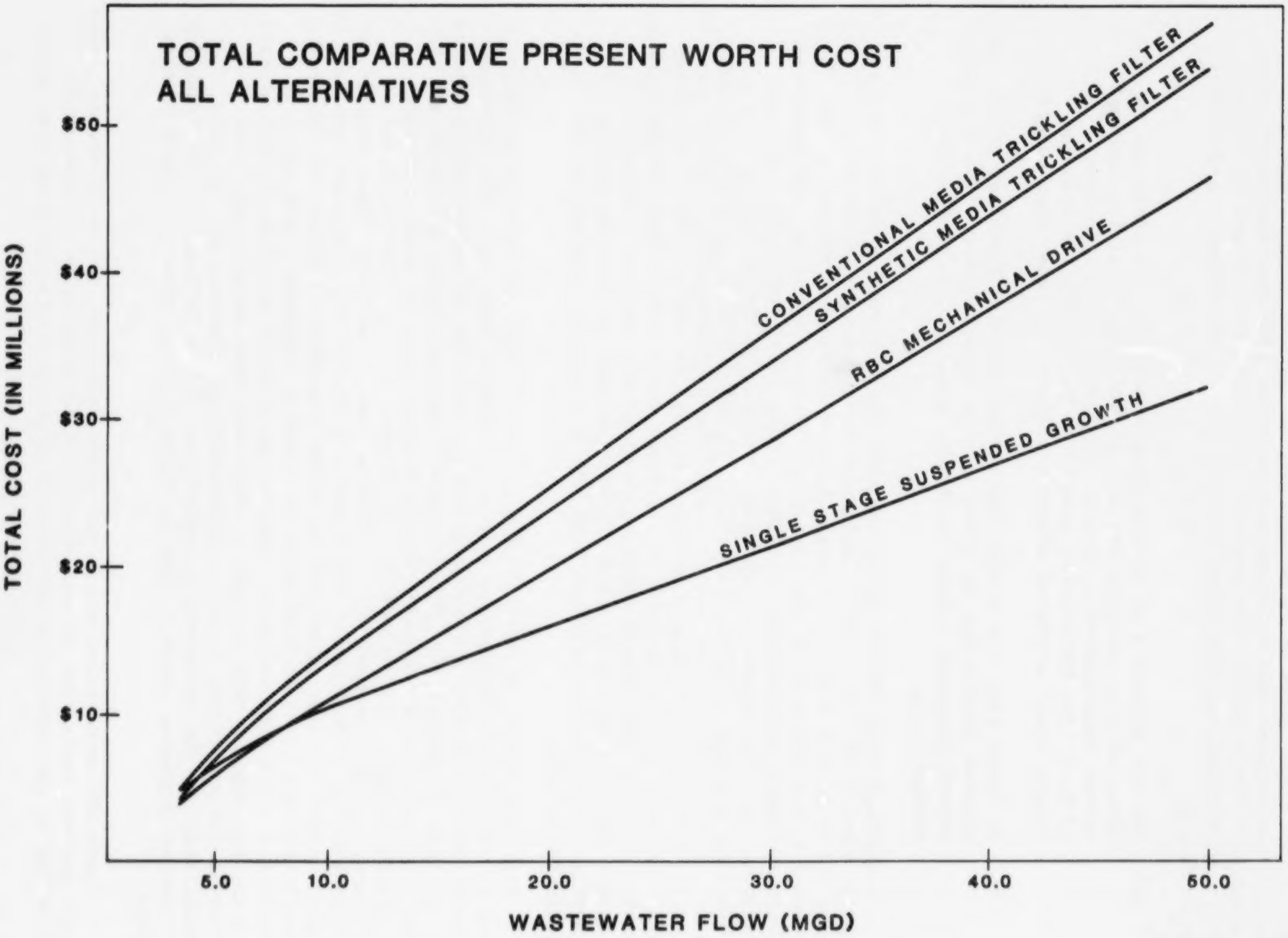

TOTAL COMPARATIVE PRESENT WORTH COST
ALL ALTERNATIVES
CONVENTIONAL MEDIA TRICKLING FILTER
SYNTHETIC MEDIA TRICKLING FILTER
RBC MECHANICAL DRIVE
SINGLE STAGE SUSPENDED GROWTH
TOTAL COST (IN MILLIONS)
$50
$40
$30
$20
$10
5.0
10.0
20.0
30.0
40.0
50.0
WASTEWATER FLOW (MGD)

FIGURE XI

It must be remembered that the above costs are not site specific and that they are developed based on an assumed set of typical design conditions. Further, preliminary cost analyses such as those developed herein are not a substitute for rigorous cost analyses conducted on a project by project basis. With these considerations in mind, only a few definite conclusions can be drawn:

A) The RBC process must be seriously considered and subjected to detailed cost analyses for all applications requiring combined carbon oxidation-nitrification involving flows less than 15 MGD. The RBC process appears to be competitive with the single stage activated sludge process to at least that capacity.

B) At flows less than 6 MGD, the synthetic media trickling filtration process should also be given serious consideration. The attached growth processes become increasingly attractive at the lower design flows.

A thorough cost-effectiveness analysis must consider more than simply present worth cost. A separate evaluation of comparative construction costs and operating costs is relevant because the lowest present worth alternative may not provide the lowest net local user charge. Alternatives involving higher construction costs but having lower operating costs may result in lower net local user charges depending on the structure of a project's financing. Figures VII, VIII, IX and X present construction cost in addition to total present worth cost. The present worth of annual operation/maintenance can be developed from these figures by subtracting construction cost from total present worth cost. Present worth annual operation/maintenance can be converted into annual operating/maintenance cost by division by the discount rate (10.49 = 20 years @ 7-1/8%) employed in this evaluation. In general, the attached growth reactors are more costly to construct but less costly to operate.

Non-monetary considerations such as reliability and operability should also be addressed in a thorough cost-effectiveness analysis. The attached growth reactors obviously have an advantage over the suspended growth process in the area of operability. They require less operational attention and are self-regulating. The suspended growth process requires close attention to sludge wastage, air demands and the recycle ratio. Such vigilence is necessary to regulate mean cell residence time, to satisfy oxygen demands and to preserve the efficiency of the final clarifiers. All processes can be called upon to provide reasonably reliable treatment. Because the attached growth reactors place less reliance on operator judgement, they could be considered to be more reliable. The design criteria for both the RBC process and the suspended growth process is reasonably well established. The criteria for both forms trickling filtration is, however, not as well established and the conservative analyst may desire to consider the trickling filtration processes somewhat less reliable than the other processes on this basis.

There are some variations to the above processes, not yet discussed, that could enhance their cost-effectiveness. As an alternative to mechanical drive RBC's, the Autotrol Corporation is offering an air drive process which offers reductions in both capital and operating cost relative to the traditional mechanical drive. In a companion Paper,[14] the Authors have shown a present worth cost savings varying from 6.5% at 3 MGD to 16% at 50 MGD relative to the traditional mechanical drive. The savings are achieved through lower power consumption and a reduction in the number of shafts of media required to reach equivalent BOD_5 and NH_3-N removals. The number of shafts are reduced due to the increased use of high density media and a slightly higher loading rate per square foot of media. It was also shown in analyses contained in that Paper that pure oxygen activated sludge was very competitive with air activated sludge at higher flows. It is expected that in the nitrification mode that pure oxygen would be even more attractive and that pure oxygen may be the suspended growth process of choice at flows over 10 MGD.

In conclusion, this study indicates that the RBC process is a viable and economic alternative for combined carbon oxidation-nitrification for wastewater flows possibly as high as 15 MGD. At higher flow rates the single stage suspended growth processes (air or oxygen activated sludge) appear to have a cost advantage. Because the RBC process has been shown to be cost-effective over a wide range of flows it should be seriously considered in future facilities planning work that involves combined carbon oxidation-nitrification.

It has also been shown that RBC design criteria has undergone a gradual evolution between 1972 and 1979. These changes have outdated information contained in fairly recently published and widely used engineering references including EPA's _Process Design Manual For Nitrogen Control_ (1975),[13] Antonie's _Fixed Biological Surfaces-Wastewater Treatment_ (1976),[15] and _Metcalf & Eddy's Wastewater Engineering: Treatment, Disposal and Reuse; 2nd Edition_ (1979).[16] The criteria contained in these texts is no longer valid and its application will result in the provision of insufficient media for nitrification and in unsatisfactory RBC performance.

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COMPUTERIZED COST EFFECTIVE ANALYSIS
OF
FIXED FILM NITRIFICATON SYSTEMS

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INTRODUCTION

Trickling filter tower systems and rotating biological contactors have been used extensively in the nitrification of municipal secondary effluents. In order to make realistic cost comparisons between these two systems, preliminary process designs have to be carried out first and capital as well as O & M costs are estimated based on these preliminary sizing calculations. Several parametric type cost curves are available in the literature which relate cost to some sizing factor; such as volume of media or wastewater flow treated (1), (2), (3) and (4). These types of curves have several disadvantages. First, they are out-dated almost as soon as they are published. Based on the authors' experience, using general cost indices to update these curves is not very successful. Second, the general cost curve approach is not sensitive enough for site-specific estimation, thus limiting the usefulness of the curves in upgrading existing treatment works.

A computerized approach may be used to accurately estimate the capital and O & M costs of both trickling filter and rotating biological contactor systems. Cost models for these processes have been developed as a part of a comprehensive computer system; CAPDET (Computer Assisted Procedures for the Design and Evaluation of Wastewater Treatment Facilities). (1)

General Description of CAPDET

The development of CAPDET is jointly funded by U.S. Army Corps of Engineers and U.S. Environmental Protection Agency. It is intended to be a screening tool capable of providing a methodology whereby a large number of alternative wastewater treatment systems, each capable of meeting specified effluent criteria, can be simultaneously ranked on the basis of cost effectiveness.

A new cost estimating approach, a "modified cost element method" is used in CAPDET. In the true cost element approach, construction details are well enough defined to adequately estimate the quantities of materials, manhours of labor, etc. necessary to build and operate the facility. Whereas, the modified cost element approach limits the detail by selecting only those cost elements which are a major impact on the total cost of the facility. Thus, total cost associated with a unit process is expressed as a sum of major component parts plus a percentage. The cost estimating procedures in CAPDET, for most unit processes, were developed on the premise that cost elements would identify at least 85% of the capital costs and 75% of the operation, maintenance and repair costs. The remainder of the costs would be expressed as a percentage of the sum of element costs.

The major cost items for construction of most unit process can generally be categorized as follows:

Earthwork
Concrete or Steel Basins
Installed Equipment
Piping
Building
Electrical and Control System
Miscellaneous Support Facilities

Detailed design formulae are developed to calculate these material quantities based on the design flow, influent pollutant concentrations, effluent limitations and user specified design parameters.

The cost of materials generally used in construction can be priced with readily available unit costs (6). However, equipment for the wastewater treatment system constitutes one of the largest items of fixed capital costs. It is desirable therefore to maintain up-to-date equipment cost data for CAPDET. With a limited number of unit cost input entries, it is very difficult to maintain reliable cost data.

The following description outlines a procedure which produces an accurate estimate of equipment cost without necessitating a vast number of unit price inputs. The installed equipment cost may be considered in three components: the purchase cost of the equipment, installation labor cost, and other minor costs, such as electrical works, minor piping, foundations, painting, etc. The purchase cost of process equipment is a function of size or capacity. To minimize the number of inputs required, a standard size (or capacity) unit is selected and the purchase cost of all other size (or capacity) units of that type is expressed as a fraction or multiple of the standard unit purchase cost. This cost ratio versus size relationship has been developed for each major item of equipment required in the design. These relationships assume the form shown:

$$\frac{(\mathrm{COST})_o}{(\mathrm{COST})_s} = F\left(\frac{A_o}{A_s}\right) \qquad (1)$$

where

A = some characteristic size measurement such as volume, area, horse power, weight.

O and S = subscripts designating other and standard sizes, respectively.

F = a function of.

The exact form of the cost-versus-size relationship and the selection of the standard sizes for each major equipment item are determined based on manufacturer's information and available literature. In most cases these size-cost relationships are relatively unaffected by inflation and other cost changes.

The CAPDET user has two options by which the purchase cost of equipment can be escalated to account for inflation. The first option is for the user to obtain from equipment manufacturers the current purchase cost of the standard size equipment. The purchase cost of any other size equipment is then automatically escalated by the cost versus size relationships which have been developed in the model. The second option is to escalate the purchase cost by the use of cost indices. Only one input is required for this process: the Marshall and Swift Equipment Cost Index from Chemical Engineering magazine. The 1977 first quarter purchase prices of the standard size equipment are stored in the model and are updated automatically if the M & S cost index is input into the program. The latter of the above methods is the least accurate, however, it provides default values when current prices are not available.

Equipment installation cost is estimated by multiplying the manhour requirements obtained in the design by user input labor rates. The other minor costs for each type of equipment consist of piping, concrete, steel, instruments, electrical, insulation, painting, insurance, taxes, etc. The cost of these items are estimated as a percentage of purchase costs and will vary with the type and size of equipment. These percentage values are established based on design experience, engineering judgement, manufacturer's input, and previously published literature.

The total construction cost of a unit process is thus the sum of the cost of installed equipment and the general construction material cost such as earthwork, reinforced concrete in place, piping, etc.

The operation and maintenance for a wastewater treatment facility can be divided into several major categories: power, operation labor, maintenance labor, chemical costs, and repair material costs.

The electrical power consumption has been determined for each unit process. The power consumption for the treatment facility is simply the totalized power consumption for the unit processes. The power consumption is converted to a cost by multiplying the power consumption in kilowatt-hours per year by the user selected unit price input for electric power. The user input unit cost should be obtained from the utility supplying power to the proposed facility.

The operation and maintenance labor costs can be divided into four groups: administration and general labor, operation labor, maintenance labor, and laboratory labor. Recommended staffing for different levels of manpower required for each of the four labor groups are established by the model. By utilizing staffing charts provided in the literature, weighted average salaries for each labor group may be established. To reduce the number of inputs, the weighted average salaries are expressed as a percent of operator II salary.

The maintenance and operation labor requirements are calculated for each individual unit process. The total requirement is the sum of the requirements for each unit process used in the treatment facility. The total annual labor cost is the sum of the labor costs for each of the four labor groups.

The cost of operation and maintenance materials and supplies is calculated for each unit process as a function of the construction cost.

The total annual operation and maintenance cost is the sum of the electric power costs, operation and maintenance labor costs, operation and maintenance material and supply costs, and chemical costs.

In order for the model to present a meaningful comparison of alternatives, a cost evaluation utilizing "time-value of money concepts" is necessary. Simply stated, the method chosen for the economic evaluation is to compare annual costs computed over a fixed evaluation period. A minimum evaluation period of 20 years is utilized in the CAPDET model. The cost evaluation has been presented such that individual unit processes or total treatment facilities may be evaluated.

Current Study

The following sections describe in detail the design and cost estimating procedures for the fixed film nitrification processes. Finally, the CAPDET model is used to compare the cost effectiveness of the two fixed film nitrification processes under various conditions.

TRICKLING FILTER NITRIFICATION SYSTEMS

The trickling filter process has been successfully utilized in a number of cases where nitrification of secondary effuent was required. Synthetic media, both low and high specific surface area types, are used in supporting the surface growth of nitrifiers. These nitrifiers have a very low yield factor, i.e., low sludge production rate; the BOD and suspended solids concentrations of the effluent are essentially the same as the influent. Therefore, final clarifiers are not generally provided in fixed film growth nitrification systems. Also, due to the light weight and high structural strength of the synthetic media, trickling filter towers can be built as tall as 28 feet without external support, thus making them economical in achieving a high degree of nitrification.

However, because of the height of modern trickling filter towers, gravity feed which is common among the old rock filters becomes a rarity. Some type of intermediate pumping is necessary to deliver the secondary effluent to the top of the filters. Therefore, a pumping station should be considered as part of the trickling filter unit process and accordingly, the associated capital and operation and maintenance costs should be included.

In this section, procedures for trickling filter tower design are outlined. The cost estimating procedures are also described. Procedures for the design and cost estimate for intermediate pumping station are discussed in the next section.

Process Design Procedures

Calculate Media Surface Area Requirement

Very few trickling filter nitrification studies have been published. Reference (7) provides the best summary of existing information. It is adopted in the CAPDET model for process calculations. The following equations can be used to calculate the required surface area based upon effluent concentration and wastewater temperature. If different concentration limitations are imposed for summer and winter conditions, the larger of the two surface area requirements is adopted as the appropriate design.

First, a critical ammonia effluent concentration is defined.

$$N_c = 4.5 - 0.115\ (T) \qquad (2)$$

where

N_c = Critical ammonia effluent concentration. mg/l as N. This gives the break points in the lines given in Reference (7).

T = Wastewater temperature, °C

The required surface area is then calculated:

If $N_e \leq N_c$

$$S = 6900 - 190\ (T) \qquad (3a)$$

If $N_e > N_c$

$$S = (21250 - 527\ T) - 3410\ N_e \qquad (3b)$$

where

S = Required surface area per pound of NH_3 - N oxidized per day, sq. ft.

N_e = Effluent ammonia concentration, mg/l as N

Calculate Volume of Media Required

$$V_d = \frac{S}{n} \times (N_o - N_e) \times 8.34 \times Q_{avg.} \qquad (4)$$

where

V_d = Volume of media required, cu. ft.

N_o = Influent ammonia concentration, mg/l as N

$Q_{avg.}$ = Averaged design wastewater flow, mgd

n = Specific surface area of the media selected. Values range from 28 to 41 sq. ft./cu. ft. depending on manufacturer's specifications

Calculate the Cross-Sectional Area and Depth of the Tower System

A surface hydraulic loading rate of 0.75 gpm/sq. ft. at design flow is recommended by the manufacturers (8) (9). The cross-sectional area then would be

$$SA = 925.3 \times Q_{avg.} \qquad (5)$$

where

SA = Cross-sectional area of filter, sq. ft.

And the depth of the tower is then calculated:

$$D = V_d/SA \qquad (6)$$

where

D = Depth or height of the media, ft.

The maximum recommended height of a tower, without external structural support, is approximately 28 feet. If the calculated depth, D, is less than 28 feet, the process design has been accomplished. A recirculation ratio of 1:1 at $Q_{avg.}$ is adequate. The pumping system should therefore be designed to handle twice the amount of the design flow or the peak flow, Q_{pk}, whichever is larger. However, under certain circumstances, it is possible that the calculated height would be larger than the maximum allowable value. An iteration procedure is provided in CAPDET to increase the tower cross-sectional area thus decreasing the height to acceptable levels. The required surface loading rate for complete wetting is provided by recirculation. These calculations would yield the design firm pumping capacity of the intermediate pumping station, Q_{FP}, expressed in mgd.

System Design

Field experience indicates that two towers are usually provided until the cross-sectional area of each tower reaches 17,700 sq. ft. or (150 ft. in diameter). When individual tower cross-sectional area, SA/2, is larger than this value, three or more towers are used. The surface area of the filter tower is limited by the available sizes of the rotary distribution arms. These arms are generally in the range of 20 to 200 feet. Figure 1 provides an experience curve which indicates the number of towers generally designed under various conditions. After the number of towers, N_t, is selected, the diameter of each individual tower can then be calculated;

$$DIA = 1.128 \left(\frac{V_d}{D \times N_t}\right)^{1/2} \qquad (7)$$

where

DIA = Diameter of individual tower, ft.

N_t = Number of towers

Estimate of Construction Costs

Construction of a typical modern trickling filter includes the media, rotating distribution arm system with its central column support, underdrain system, and the outside wall. The plastic media usually is supplied and installed by the manufacturers and the cost of the installed system is estimated as dollars/ cu. ft. Polyester fiberglass, light weight steel, and precast double-tee concrete construction have all been used as the media containment structure. In CAPDET, study, it is assumed that a 6-inch cast-in-place reinforced concrete wall is utilized. The media support system consists of precast beams and concrete support posts as suggested by one of the manufacturers (9). The underdrain system includes the drainage floor and channel, sidewall with air openings and louvers. The distributor system is supported by a concrete column extending from the floor to the top of the tower. The following procedures are designed to estimate the material takeoff of these major items and subsequently, the construction cost by using unit price inputs.

Material Takeoff

The major cost items in a trickling filter system include cost of earthwork, cost of reinforced concrete in place, cost of medium and cost of distributor arm. Table 1 summarizes the algorithms for calculating the quantities of material required. These quantities are related to the specific dimensions of a selected tower.

Cost of Equipment

Two major items are involved here, the cost of the media and the cost of the distributor arms. The cost of the plastic sheet media ranges from \$2.50 to \$4.00 per cu. ft. depending on the specific surface area. The estimate of the cost of the media, $\$_{media}$, is straight forward

$$\$_{media} = 0.785 \times (DIA)^2 \times D \times N_t \times UPIMC \qquad (8)$$

where

UPIMC = Media unit cost, a user input in dollars per cu. ft.

Calculating the cost of distributor arm system is more complicated. The purchase cost of this type of equipment can be related to the cost of a standard size arm, in this case, a distributor arm with a diameter of 50 ft.

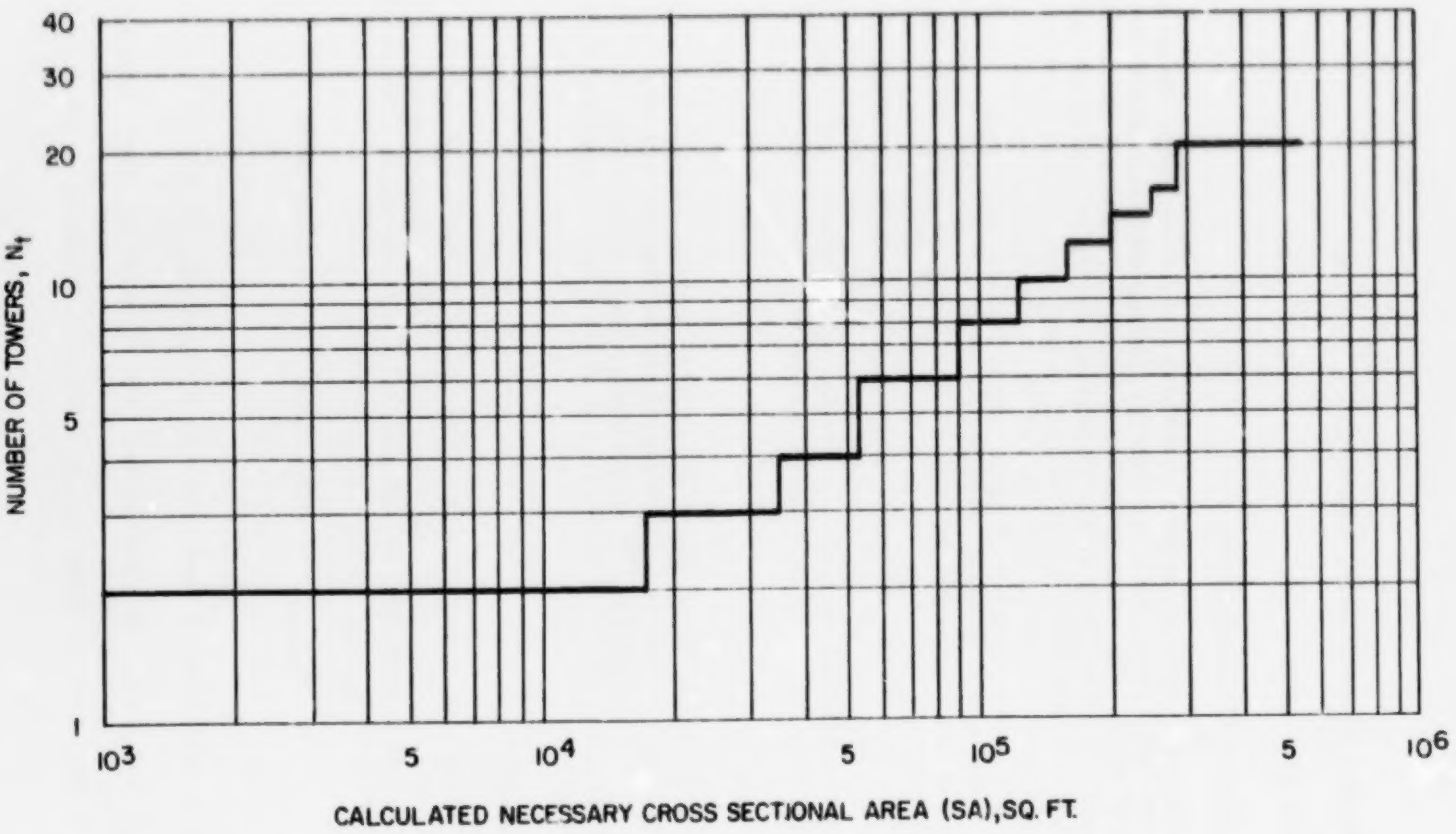

FIGURE I

PROCEDURES FOR THE SELECTION OF TRICKLING FILTER TOWERS

Table 1

Quantities of Material for the Construction of a Trickling Filter System

I. Earthwork Quantities, V_{ew}, (cu. yd.)

$$V_{ew} = 1.48 \times 10^{-3} (DIA)^3 + 0.208 (DIA)^2 + 3.28 (DIA) + 14.9$$

II. Reinforced Concrete Wall, V_{cw}, (cu. yd.)

i. External wall

$$V_{cwe} = 0.0581 (D + 3) (DIA) + 0.467 (DIA)$$

ii. Media support system

If DIA $\leq$ 40 ft.

$$V_{cws} = 5.33 \times 10^{-5} (DIA)^{3.08} + 7.41 \times 10^{-4} (DIA)^{2.963}$$

If DIA $>$ 40 ft.

$$V_{cws} = 3.64 \times 10^{-3} (DIA)^{1.936} + 0.023 (DIA)^{2.035}$$

iii. $V_{cw} = V_{cwe} + V_{cws}$

III. Reinforced Concrete Slab, V_{cs}, (cu. yd.)

If DIA $\leq$ 70 ft.

$$V_{cs} = 0.019 (DIA)^2 + 0.37 (DIA) + 0.151 D + 0.3$$

If DIA $>$ 70 ft.

$$V_{cs} = 0.019 (DIA)^2 + 0.63 (DIA) + 0.6 D + 1.2$$

V_{ew} = Volume of earthwork, cu. yd.

V_{cw} = Volume of reinforced concrete wall in place, cu. yd.

V_{cs} = Volume of reinforced concrete slab in place, cu. yd.

The relationship is;

$$\$_{Arm} = \left[0.367 + 0.01265\ (DIA)\right] \times \$_{Arm - 50} \qquad (9)$$

where

$\$_{Arm}$ = The purchase price of a distributor arm with a diameter of DIA feet, dollars

$\$_{Arm - 50}$ = The standard size arm cost to be specified by the user, dollars

Construction Cost

The construction cost of the tower system: minus the usual miscellaneous costs such as contractor's profit, engineering fee, general site work and others, can be estimated by using standard construction unit costs. The relationship is:

$$\$_{TF} = 1.10 \left[\$_{media} + N_t\ (1.32\ \$_{Arm} + \$_{ew} + \$_{cw} + \$_{cs})\right] \qquad (10)$$

where

$\$_{TF}$ = Construction cost of trickling filter systems, dollars

1.10 = Cost increase accounting for the minor items such as piping, stairway and etc.

1.32 = Cost escalation accounting for the installation of the arm system

and

$$\$_{ew} = V_{ew} \times UPIEW \qquad (11a)$$

$$\$_{cw} = V_{cw} \times UPICW \qquad (11b)$$

$$\$_{cs} = V_{cs} \times UPICS \qquad (11c)$$

UPIEW = Unit price input of earthwork, dollars per cu. yd.

UPICW = Unit price input of reinforced concrete wall in place, dollars per cu. yd.

UPICS = Unit price input of reinforced concrete slab in place, dollars per cu. yd.

Estimate Operation and Maintenance Cost

The operation and maintenance (O & M) cost of running a trickling filter system includes the cost of manpower and that of repair and maintenance material. The tower itself does not require any significant amount of energy. Although in certain areas of the U.S., towers are equipped with mechanical ventilation devices, the energy required to operate these fans is minimal. Thus, the only energy required to operate the trickling filter nitrification process would be pumping which will be described in the next section.

O & M Manpower

Reference (1) supplies O & M manpower requirements with respect to the surface area of the towers:

If SA $\leq$ 11,000 sq. ft.

$$OMH = 10\ (SA)^{0.30} \tag{12}$$

and

$$MMH = 11.5\ (SA)^{0.26} \tag{13}$$

If SA $>$ 11,000 sq. ft.

$$OMH = 0.2 \times (SA)^{0.7} \tag{14}$$

$$MMH = 0.046 \times (SA)^{0.86} \tag{15}$$

The total operation and maintenance manpower requirements would be:

$$OMMH = OMH + MMH \tag{16}$$

where

OMH = Operation manpower requirement, manhrs/yr.

MMH = Maintenance manpower requirement, manhrs/yr.

OMMH = O & M manpower requirement, manhrs/yr.

SA = Cross-sectional area of towers, sq. ft.

Operation and Maintenance Material Cost

This item includes repair and replacement material cost and cost of major maintenance work performed by outside contractors. It is approximated by a percentage of the construction cost of the tower system (1).

$$\$_{MR} = 0.006 \times \$_{TF} \tag{17}$$

where

$\$_{MR}$ = Annual maintenance and repair material cost

Total O & M Cost

The total O & M cost can be easily estimated by using a unit price input for the labor rate.

$$\$_{om} = (OMMH)(MC) + \$_{MR} \tag{18}$$

where

$\$_{om}$ = Annual O & M cost, dollars/yr.

MC = Manpower cost, dollars/manhr.

INTERMEDIATE PUMPING STATIONS

An intermediate pumping station is an integral part of a modern trickling filter system. The secondary effluent from the final clarifier will flow into a wet well and then be lifted by pumps to the top of the tower. The pumps not only have to deliver the wastewater volume but also to create enough head to cause the distributor arms to rotate. Ordinarily, constant speed centrifugal pumps are employed. The logic for selection of a constant speed pump is that a more or less constant hydraulic loading to the towers is always maintained in field practice.

The intermediate pumping station includes the pumps and their associated piping and controls, the wet well, dry well and motor housing. The preliminary design and cost estimating procedures are described briefly here.

Process Design Procedures

In pumping system design, two parameters will have to be defined first; design flow or firm pumping capacity and the total dynamic head. The design firm pumping capacity, Q_{FP}, has been specified by the logic in the section on trickling filter.

The total dynamic head of this pumping system would include (1) the static head, (2) the frictional head created by flow velocity through the pipes and (3) the dynamic head required to rotate the distributor arms. Experience has shown that the following equation gives a reasonable prediction for various flow and site conditions:

$$TDH = D + 12 \tag{19}$$

where

TDH = Total dynamic head, ft.

D = Height of the filter media, ft.

Pump Selection

In selecting the number of pumps within the pump station, the rule-of-thumb summarized in the following will be used.

Selection of Number of Pumps Within
a Pump Station

Q_{FP}, (mgd)	Number of Pumps, N_p
Less than 5 mgd	2
5 - 50 mgd	3
50 - 150 mgd	4
more than 150 mgd	5

The number selected, N_p, includes one standby pump. For simplicity, the same size pumps are used in this model, then each individual pump would handle:

$$q = 694.4 \frac{Q_{FP}}{N_p - 1} \qquad (20)$$

where

q = Capacity of each pump, gpm

It is customary to add a safety factor to the calculated pump capacity. Values from 10 to 50 percent have been suggested (10). The following equation gives a relationship between the suggested safety factor and the sizes of pumps.

$$SF_p = 1.9 - 0.2 \text{ Log } q \qquad (21)$$

and SF_p is always larger or equal to 1.1

where

SF_p = Safety factor for pump selection

Thus, the design flow for each pump would be

$$q_d = (SF_p) \times (q) \qquad (22)$$

where

q_d = Design capacity of the selected pump, gpm

Experience has shown that the pump discharge nozzle velocity should be in the range of 11 to 14 feet per second for wastewater applications. And the sizing procedure would be to calculate the pump's discharge nozzle diameter.

$$d_p = 0.1927 \quad q_d \qquad (23)$$

d_p = Pump sizes in inches of discharge nozzle diameter. The available pump sizes are 4", 6", 8", 10", 12", 14", 16", 18", 20", 24", 30", 36", 42", 48", 54", 60" and 72". A step function is employed in CAPDET to select the appropriate pump from the available sizes.

Drive Selection

Electric motors are the most commonly employed pump drivers in a wastewater treatment plant. The cost of a motor is related to its rated horsepower and rotating speed. The rotating speed is usually governed by the characteristics of the pumps selected. An important index of centrifugal pumps is the specific speed parameter. It is considered as a shape factor.

The specific speed is defined as (10):

$$NS = \frac{r \times (q_d)^{0.5}}{(TDH)^{0.75}} \tag{24}$$

where

NS = Specific speed

r = Impeller rotating speed, rpm

A typical value of 4000 has shown to be most appropriate for sewage applications (10). Thus the rotating speed for the motor would be:

$$r = \frac{4000\ (TDH)^{0.75}}{(q_d)^{0.5}} \tag{25}$$

On the other hand, the brake horsepower to be delivered by the motor is directly related to the flow and the head;

$$BH_p = \frac{(q_d)(TDH)}{3959.7\ (E)} \tag{26}$$

where

BH_p = Brake horsepower of the motor, H_p

E = Pump efficiency

Pump efficiency is a variable changing with the head it is pumping against. However, it is a rule that the larger the pump the more efficient it is. And a general relationship has been developed as follows (12):

$$E = 0.63 + 2.42 \times 10^{-5} \times (q_d) \tag{27}$$

and E is always less than or equal to 0.90

Again, a step function is used in CAPDET to select the motor from commercially available sizes.

Estimate of Construction Costs

The major construction items of an intermediate pumping station, besides the cost of pumps and equipment, consists of the wet well, dry well and motor control housing. The wet well and dry well are usually underground structures which can be priced by estimating the quantities of earthwork and reinforced concrete required. The cost of the building housing the motors and their control panels can be approximated by multiplying a unit cost to the square footage of the building.

Material Takeoff

The general algorithms for calculating material quantities for the pumping station construction is summarized in Table 3. The relationships are developed from field data. They are related to the firm pumping capacity, Q_{FP}.

Cost of Equipment

Two major cost items are to be developed for each pump station; the cost of pumps and the cost of the drive units. The other costs such as piping, control, ventilation, dehumidification can be substantial. However, historically, the "minor" costs have been estimated as a percent of the cost of pump and driver.

The purchase price of a single pump is a nonlinear function of the pump size expressed as the diameter of the discharge nozzle. The purchase cost of pumps can be related to the cost of a standard size pump and 16 inch pump has been found to be the most appropriate. The relationship is shown as follows: (12)

If $d_p \leq 16$ inches

$$\$_p = 0.041\ (d_p)^{1.152} \times (\$_{p\ -\ 16}) \qquad (28a)$$

If $d_p > 16$ inches

$$\$_p = 0.01\ (d_p)^{1.663} \times (\$_{p\ -\ 16}) \qquad (28b)$$

where

d_p = Pump size, inches

$\$_p$ = Purchase price of pump at size d_p, dollars

$\$_{p\ -\ 16}$ = Purchase price of a 16 inch-ø pump, dollars

The purchase price of an AC motor is a function of the power it can deliver, BH_p, and its synchronous speed. Again, the price of any motor can be related to that of the standard. In this case, it is an induction motor with 100 Hp capacity and 1200 rpm speed. The relationship is shown as: (13)

$$\$_m = 3.9\ \frac{(BH_p + 5)^{0.85}}{r^{0.75}}\ (\$_{m\ -\ 100}) \qquad (29)$$

where

$\$_m$ = Purchase price of motor, dollars

$\$_{m\ -\ 100}$ = Purchase price of the standard motor, dollars

Table 3

Quantities of Material for the Construction of an Intermediate Pumping Station

I. Earthwork Quantities, V_{ew}, (cu. yd.)

If $Q_{FP} \leq 20$ mgd

$$V_{ew} = 1{,}519\ (Q_{FP})^{0.22}$$

If $Q_{FP} > 20$ mgd

$$V_{ew} = 591\ (Q_{FP})^{0.538}$$

II. Reinforced Concrete Wall, V_{cw}, (cu. yd.)

If $Q_{FP} \leq 10$ mgd

$$V_{cw} = 77.8\ (Q_{FP})^{0.147}$$

If $Q_{FP} > 10$ mgd

$$V_{cw} = 50.3\ (Q_{FP})^{0.337}$$

III. Reinforced Concrete Slab, V_{cs}, (cu. yd.)

$$V_{cs} = 80.4\ (Q_{FP})^{0.408}$$

IV. Pump Station Building Area, A_b, (sq. ft.)

$$A_b = 19.7\ (Q_{FP}) + 640$$

Guthrie (14) gives the following factors to be used in estimating field installation cost for pumps and their drive units;

Items	Percent of Equipment Cost
Equipment Cost (pump and motor)	100.0 %
Field Materials	
Piping	29.6 %
Concrete	3.9 %
Instruments	2.9 %
Electrical	30.3 %
Insulation	2.5 %
Paint	0.8 %
Field Labor	
Material Erection	59.0 %
Equipment Setting	8.9 %
Total Direct Cost	237.9 %

Construction Cost

The construction cost of an intermediate pumping station can thus be estimated by:

$$\$_{PS} = 1.15 \times \left[N_p \times (2.38)(\$_p + \$_m) + \$_{ew} + \$_{cw} + \$_{cs} + \$_b \right] \tag{30}$$

where

$\$_{PS}$ = Construction cost of intermediate pumping station, dollars

1.15 = Cost increases accounting for the minor articles such as overhead crane, hvac, etc.

and

$$\$_b = A_b \times \text{UPIBC} \tag{31}$$

Where UPIBC is the unit price for building expressed in dollars per square feet. Other symbols have been defined earlier.

Estimate of Operation and Maintenance Costs

The major items of the intermediate pump station O & M cost are: O & M manpower requirement, energy and material and supply cost. They are estimated individually in the following subsections.

O & M Manpower

The available information relates the O & M manpower requirement with the fir pumping capacity of the station (3).

If $0 < Q_{FP} \leq 7$ mgd

$$OMH = 440\ Q_{FP}^{0.129} \tag{32}$$

$$MMH = 360\ Q_{FP}^{0.148} \tag{33}$$

If $7 < Q_{FP} \leq 30$ mgd

$$OMH = 294\ Q_{FP}^{0.335} \tag{34}$$

$$MMH = 255\ Q_{FP}^{0.325} \tag{35}$$

If $30 < Q_{FP} \leq 80$ mgd

$$OMH = 40\ Q_{FP}^{0.866} \tag{36}$$

$$MMH = 86\ Q_{FP}^{0.646} \tag{37}$$

If $Q_{FP} > 80$ mgd

$$OMH = 21\ Q_{FP}^{1.012} \tag{38}$$

$$MMH = 31\ Q_{FP}^{0.881} \tag{39}$$

where

OMH = Operation manpower requirement, manhrs/yr.

MMH = Maintenance manpower requirement, manhrs/yr.

Q_{FP} = Firm pumping capacity, mgd

Total manpower requirement OMMH would be

$$OMMH = OMH + MMH \tag{40}$$

Annual Power Consumption

As mentioned previously, essentially the only power requirement for the trickling filter system is the pumping process. The usual operating procedure is to keep the hydraulic loading rate around 0.75 gpm/sq. ft. Thus, the operating flow would be:

$$q_0 = 0.589\ (N_t)(DIA)^2 \tag{41}$$

where

q_o = Operation flow, gpm

The pumping energy would be:

$$KWH = 1.66 \times \frac{(q_o)(TDH)}{E} \tag{42}$$

where

KWH = Annual power requirement, kwhr/yr.

Operation and Maintenance Material Cost

This item includes repair and replacement material cost and the cost of major maintenance works to be performed by outside contractors. It can be expressed as a percentage of the construction cost of the intermediate pump station (1).

$$\$_{MR} = 0.01 \ \$_{ps} \tag{43}$$

where

$\$_{MR}$ = Cost of maintenance and repair material, dollars/yr.

Total O & M Cost

$$\$_{om} = (OMMH)(MC) + (KWH)(EC) + \$_{MR} \tag{44}$$

ROTATING BIOLOGICAL CONTACTOR NITRIFICATION SYSTEM

The RBC is a relatively new process. It has also been utilized in promoting nitrification on secondary effluents.

The basic principles governing the performance of the RBC systems are the same as those for trickling filters. The only major process difference between these two systems is that for the RBC process the aeration and mass transfer mechanisms are provided by rotating the media in and out of the wastewaters, whereas for trickling filter systems, aeration and mass transfer mechanisms are provided by passing the wastewater over the media.

The following subsections provide the design and cost estimating procedures for the RBC system.

Process Design Procedures

Figure 2 shows the experimental relationships between surface nitrification rate and the exposed ammonia concentration for effluents (15). As one can observe, the rate is related to the temperature and the effluent ammonia concentration desired. The information presented in Figure 4 can be modeled by using the same technique proposed by Keinath (16). The resulting design equation is as follows:

$$A_N = \frac{15,125}{TCF} \left[8.6 \ln \left(\frac{N_o}{N_e}\right) + (N_o - N_e) \right] \tag{45}$$

where

A_N = Surface area required per million gallons of wastewater flow, sq. ft./mgd

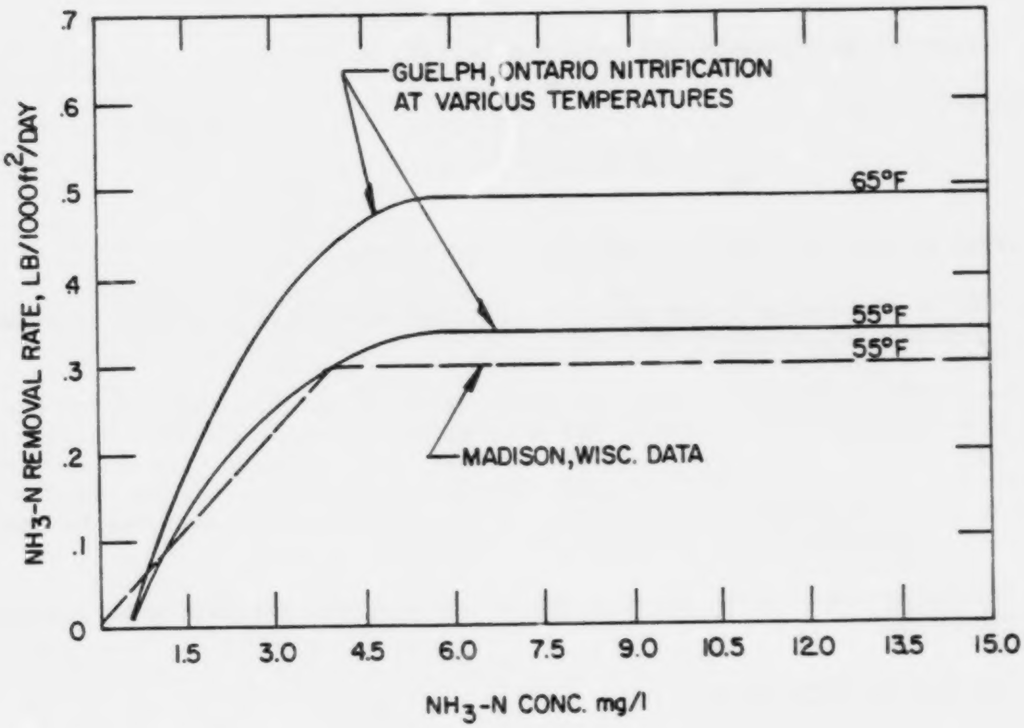

FIGURE 2

AMMONIA REMOVAL RATE
FOR RBC SYSTEMS

TCF = Temperature correction factor

and TCF, Temperature Correction Factor can be defined as:

When T is larger than 18° C

$$TCF = 1.4 \tag{46a}$$

When T is smaller than 18° C

$$TCF = 0.0787 \times T \tag{46b}$$

If different effluent concentration limitations are given for summer and winter conditions, the surface area requirement for each condition must be determined and the final design based on the condition that requires a larger media surface.

The required media surface area would be:

$$A = Q_{avg.} \times A_N \tag{47}$$

where

A = Required media surface area, sq. ft.

Estimate of Construction Costs

Working from the required media surface area, a physical layout of the plant is assumed from which construction costs, such as excavation, equipment and concrete, are derived.

The first step in determining capital costs is to calculate the number of RBC shafts required to supply the design media surface area. Since generally a large amount of surface area is needed to support the less efficient nitrifying biomass, shafts having areas of 150,000 ft.^{2} each are commonly used. This shaft has physical dimension of 11'-10" in diameter and 25' in length. The specific surface area is approximately 54.5 sq.ft./cu.ft. (17) The number of shafts required may be calculated as follows:

$$N_{sh} = \frac{A}{A_{sh}} \tag{48}$$

where

N_{sh} = Number of shafts necessary

A = Total media surface area required, sq. ft.

A_{sh} = Media surface area per shaft, sq.ft./shaft, 150,000 sq.ft./shaft

For practical reasons the value for N_{sh} is rounded up to the next whole number to avoid working with fractional shafts in subsequent calculations.

Material Takeoff

For simplicity of design, it is assumed that shafts will be arranged in groups of eight. Each group is called a bank and consists of two end shafts and six intermediate shafts. Any shafts in excess of a multiple number of eight are placed in a partial bank of from one to seven shafts as needed. Many other configurations are possible; however, varying the configurations does not affect earthwork, concrete, and other related material requirements significantly. The number of full and partial banks, N_b, is, therefore, the number of shafts divided by eight and rounded up to the next whole number.

$$N_b = N_{sh}/8 \qquad (49)$$

where

N_b = Number of banks

Each bank will consist of two end shafts and from zero to six intermediate shafts. The number of end shafts, N_{es}, and intermediate shafts, N_{is}, are calculated as follows:

$$N_{es} = 2\ N_b \qquad (50)$$

$$N_{is} = N_{sh} - N_{es} \qquad (51)$$

Using these figures and the associated volumes they represent, the earthwork and concrete requirements are calculated.

$$V_{ew} = 130\ N_{es} + 142\ N_{is} \qquad (52)$$

$$V_{cs} = 23\ N_{es} + 20.7\ N_{is} \qquad (53)$$

$$V_{cw} = 11.5\ N_{es} + 8.6\ N_{is} \qquad (54)$$

where

V_{ew} = Volume of earthwork required, cubic yards

V_{cs} = Volume of concrete slab required, cubic yards

V_{cw} = Volume of concrete wall required, cubic yards

Construction Cost

The bare construction cost of the RBC system consists of the costs of earthwork, concrete works, installed media and other minor items. The RBC shafts are purchased as a complete unit from the manufacturers. Usually the package includes the media, protective fiberglass cover, driving unit and other miscellaneous items. A factor of 15% is added to the RBC media purchase cost for installation.

The construction cost of a RBC system would therefore be:

$$\$_{RBC} = 1.10 \left[1.15\ (N_{sh}) \times \$_{sh} + \$_{ew} + \$_{cw} + \$_{cs} \right] \qquad (55)$$

where

$\$_{RBC}$ = Construction cost of RBC system, dollars

1.10 = Cost increase accounting for minor items such as site cleaning, piping, electrical work and etc.

Estimate of Operation and Maintenance Costs

The major components of the RBC operation and maintenance cost are the O & M manpower requirement, energy and material and supply cost.

O & M Manpower

Reference (17) provides an estimate of manpower for the operation and maintenance of the RBC system;

For a plant with less than 30 shafts

$$OMMH = 52\ N_{sp}\ (1.25 - 0.025\ N_{sp}) \qquad (56a)$$

For a plant with more than 30 shafts

$$OMMH = 26 \times N_{sh} \qquad (56b)$$

where

OMMH = Operation and maintenance manpower, manhrs/year

Annual Power Consumption

A power consumption of 5 Hp/shaft is used in estimating the power consumption of the RBC system. (18)

$$KWH = 32{,}630 \times N_{sh} \qquad (57)$$

where

KWH = Annual power consumption, kwh/yr.

Operation and Maintenance Material Cost

This item includes repair and replacement material cost and the cost of major maintenance works performed by outside contractors. It is expressed as a percent of installed cost of the RBC equipment.

$$\$_{MR} = 0.01\ (1.15)(N_{sh})\ \$_{sh} \qquad (58)$$

Total O & M Cost

$$\$_{om} = (OMMH)(MC) + (KWH)(EC) + \$_{MR} \quad (59)$$

where

$\$_{om}$ = Annual O & M cost, dollars/yr.

MC = Manpower cost, dollars/manhr.

EC = Electricity cost, dollars/kwhr

COST EVALUATION OF FIXED FILM NITRIFICATION SYSTEMS

The CAPDET model was used to perform the comparative cost evaluation of the two fixed film nitrification systems described above. The inputs used for the cost evaluation are summarized in Table 4. The analysis was performed for wastewater flows ranging from 0.5 to 50 mgd. The CAPDET model allows change of any or all of the parameters presented by Table 4 in order to reflect local design conditions. However, space limits the quantity of data which can be presented in this paper. In addition to the design factors used, the assumption is also made that the peak flow is twice the average flow in design of the intermediate pumping station associated with the trickling filter tower.

The results of the cost analysis conducted during this study are presented graphically by Figures 3 and 4. As stated, this study was designed to evaluate the affects of increasing flows on the relative cost effectiveness of the two fixed film nitrification systems.

The values presented in these figures represent the bare construction, operation and maintenance costs of the individual unit processes. Contractor profit and overhead, mobilization and sitework as well as engineering costs are not included.

Under the conditions presented in Table 4, the RBC system is more cost effective than the trickling filter system when the design flow is less than 3 mgd. However, for larger plants the trickling filter system becomes more competitive. It is hypothesized that because of the modular nature of the RBC system, its "economy of scale" is not as favorable as that of the trickling filter towers. However, the air drived RBC system might change the cost picture significantly. Another point that has to be emphasized here is that the process design equations for trickling filter nitrification systems were based on pilot studies under controlled environment. No full scale plant performance data is available to substantiate their accuracy. Whereas, the design equations for RBC was derived from full scale plant studies. It is highly possible that the surface areas requirement for tickling filter may have to be increased based upon future findings. The CAPDET model can be easily upgraded to accommodate the constant expanding knowledge of the process kinetics of fixed film nitrification processes.

Table 4

Conditions Sepcified in the Cost Evaluation of the Fixed Film Nitrification Systems

I. Influent, Effluent Conditions

- i. Influent oxidizable nitrogen concentration: 25 mg/l-N
- ii. Effluent ammonia limitation: summer 1 mg/l-N @ 24°C
  winter 4 mg/l-N @ 9°C

II. Trickling Filter Nitrification System

- i. Media cost: UPIMC = $3.50/cu.ft. @ 41 sq.ft./cu.ft.
- ii. Standard size distributor arm cost: $\$_{Arm - 50}$ = $39,000

III. Intermediate Pumping Station

- i. Standard size pump cost: $\$_{p - 16}$ = $16,500
- ii. Standard size motor cost: $\$_{m - 10}$ = $2,850
- iii. Pump house building cost: UPIBC = $30/sq. ft.

IV. RBC Nitrification System

- i. RBC medium unit cost: $39,000 @ 150,000 sq.ft./shaft

V. General Civil Works

- i. Unit price of earthwork: UPIEW = $5.00/cu.yd.
- ii. Unit price of reinforced concrete wall: UPICW = $300/cu.yd.
- iii. Unit price of reinforced concrete slab: UPICS = $270/cu.yd.

VI. Operation and Maintenance Unit Costs

- i. Labor rate: MC = $7.00/manhr.
- ii. Cost of electricity: EC = $0.04/kwhr.

VII. Parameters Used In Economic Analysis

- i. Interest rate: 6 7/8% annual
- ii. Return period: 20 years

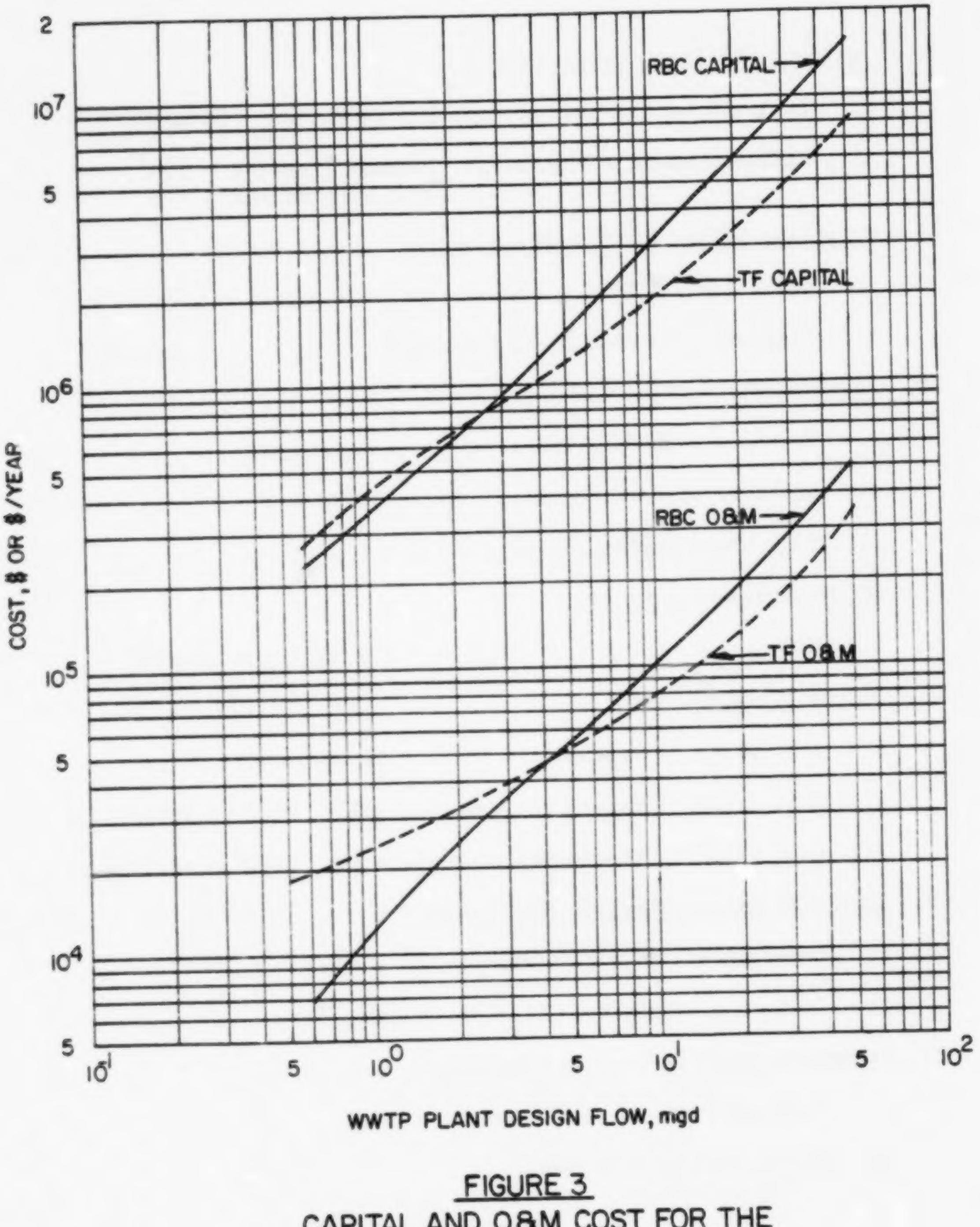

FIGURE 3

CAPITAL AND O&M COST FOR THE FIXED FILM NITRIFICATION SYSTEMS

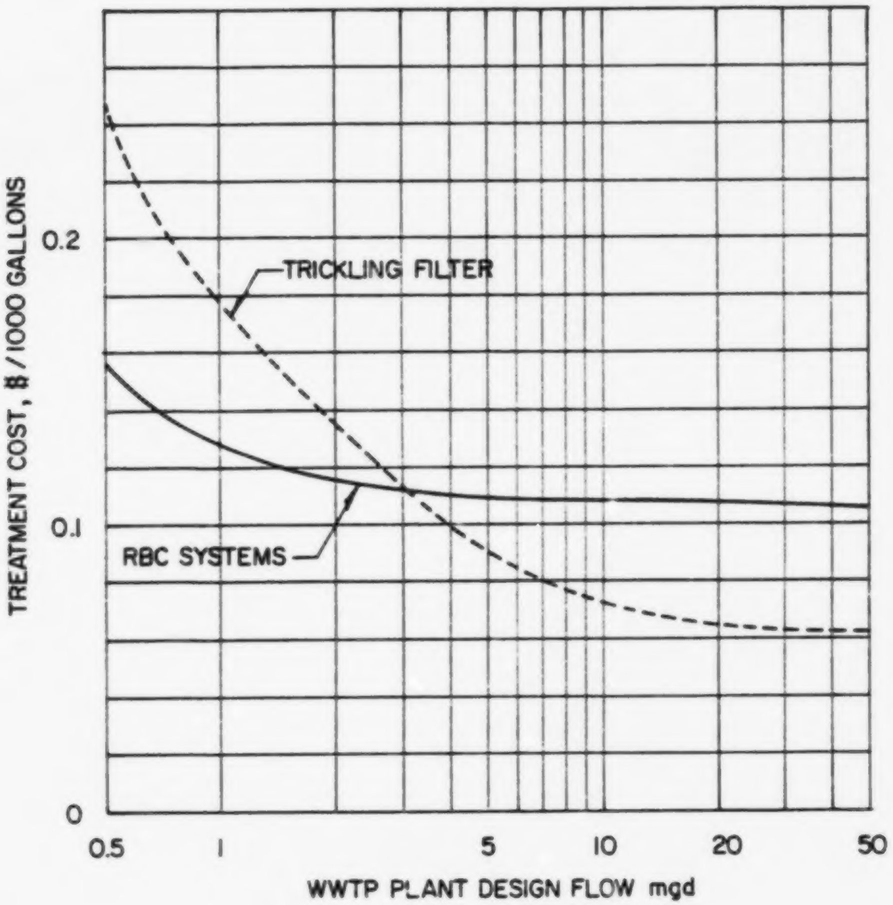

FIGURE 4

COMPARISON OF TREATMENT COSTS OF THE TWO FIXED FILM NITRIFICATION SYSTEMS

SUMMARY

A detailed description of the CAPDET model for the fixed film nitrification systems is presented here. The model is so designed with many flexibilities are incorporated. Engineers can obtain an accurate cost analysis with few user inputs. They can manipulate the unit price inputs to suit the local conditions, if desired. For instance, for sites with difficult soil characteristics, a high unit price value for earthwork can be used, thus, yielding more realistic cost comparisons. CAPDET is also a valuable research and education tool. It can provide students the opportunity to study the cost savings of certain process improvement and the optimization of overall treatment plant design.

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COMPARATIVE COST-EFFECTIVENESS ANALYSIS OF ROTATING BIOLOGICAL CONTACTOR AND ACTIVATED SLUDGE PROCESSES FOR CARBON OXIDATION

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INTRODUCTION

The use of cost-effectiveness analysis as a decision-making tool in facilities planning for wastewater treatment works is required under the U.S. Environmental Protection Agency's regulations governing its construction grants program as currently promulgated under the Clean Water Act of 1977 (P.L. 95-317). This tool has been applied routinely for municipal treatment works since the original regulations were issued pursuant to the Federal Water Pollution Control Act of 1972 (P.L. 92-500). When properly applied, it can provide a powerful insight into the total costs associated with competing alternative courses of action and, when expanded upon, can yield a system formula which may be analyzed for sensitivity to differential inflation factors between the resources required for each alternative. Unfortunately, the various methods historically used for cost-effectiveness analysis have been inconsistent throughout the engineering community due to different interpretations of the regulatory guidelines; therefore, the resultant alternative selection process has tended to be, in many cases, more subjective than objective in nature. In certain instances, these inconsistencies may have led to either the premature rejection of the Rotating Biological Contactor (RBC) process from consideration at the screening level or the weighting of the cost-effectiveness analysis so as to place the RBC

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process at a competitive disadvantage against other more "conventional" wastewater treatment processes.

The purpose of this study is to present a summary of the results of cost-effectiveness analyses which compare air-drive and mechanical-drive RBC processes with air and pure oxygen activated sludge processes over a range of design flow capacities. The general methodology used in the analyses is developed in such a manner as to reduce the impact of subjective variables on the analyses and to provide a clearer picture of the relative merits of each alternative process. The examples selected involve the upgrading of an existing primary treatment plant to meet secondary effluent limitations of 30 mg/l BOD_5 and 30 mg/l suspended solids. Typical design parameters are used for the development of preliminary cost estimates for each generic process.

It is important to note that the cost estimates set forth herein are not meant to represent optimized designs on a case by case basis. For example purposes, sufficient information has been assembled to enable the preparation of preliminary estimates which should reflect the relative rankings of the competing processes. It is recognized that every analyst has his own estimating methods and, further, that any specific project has its own idiosyncracies; however, it is the author's contention that, when comparative estimates are made using the same basic procedures and assumptions for each alternative, the relative costs of the various processes will be appropriate for the purpose of determining the most cost-effective alternative(s).

NOTES ON COST-EFFECTIVENESS ANALYSIS

Performing a cost-effectiveness analysis is likely the most important step in the facilities planning process. Viewed in a simplistic manner, facilities planning consists of four steps typical of most engineering problems, namely: problem definition, data collection, analysis of data and formulation of conclusions. While a certain degree of flexibility exists with regard to the selection of the final alternative on the basis of subjective judgement, one generally has great difficulty in justifying the selection of an alternative other than the one with the lowest present worth unless there are several other alternatives which rank very closely to it in terms of their relative present worth. For this reason, some analyses may have been prepared in which certain alternatives were unfairly weighted or eliminated at the screening level when, in fact, they may have been the most cost-effective if the analysis had been conducted more objectively. Being comparatively young in contrast to activated sludge processes, RBC processes may still be subject to significant reservations within the engineering community and may have fallen prey to subjective screening in past facilities planning efforts. This is probably not as prevalent today, as RBC processes have become much more widely accepted in recent years. This is evidenced by the large number of RBC plants currently in operation, under construction or being designed. Further, the influx of several new RBC manufacturers into the marketplace within the past few years provides additional evidence of their widespread acceptance.

In order for a cost-effectiveness analysis to be truly useful, it must include all potential process chains which are capable of meeting the required effluent standards and likely to be cost-competitive. Once the competing process chains have been selected, the analyst must collect and/or develop sufficient information to enable him to estimate, as accurately as possible, the complete economic impacts which would result from the selection of any one of the competing alternatives. All process-dependent facilities should be included and all non process-dependent facilities such as laboratory buildings, maintenance garages, administrative buildings, etc. should, for the purposes of analysis, be excluded, unless the costs for these facilities could change as a result of the process which is selected. The degree to which the various process details are analyzed generally will not seriously affect the ultimate outcome of the analysis, provided that at least a minimum level of detail is used and, further, that all competing alternatives are developed to the same level of detail. One should avoid the practice of developing greatly detailed estimates for certain processes, including careful enumeration of equipment, facilities, operation and maintenance requirements and construction costs (the "quasi-design" approach) while, at the same time, developing cost estimates for the remaining competing processes based on published cost curves (the "pseudo-design" approach). The effect of intermingling approaches at this juncture in the facilities planning process could invalidate the results generated by the analysis and, as a consequence, render possibly questionable ultimate conclusions. Following completion of the cost-effectiveness analysis, subjective considerations may be addressed again, if necessary, as in the case where the analysis has yielded more than one alternative with the lowest present worth or very close values, to within the accuracy of the estimates.

BASIS FOR EXAMPLES

To demonstrate the potential range of design flow capacities over which RBC processes may be cost-competitive, sample cost-effectiveness analyses have been prepared for the upgrading of hypothetically existing primary treatment plants to meet secondary effluent limitations of 30 mg/l BOD_5 and 30 mg/l suspended solids for design flow capacities of 3, 5, 10, 30 and 50 MGD. The influent to the secondary treatment facilities from the existing primary portion of the plant has been assumed to have the following characteristics:

BOD_5 (Total):	140 mg/l
BOD_5 (Soluble):	85 mg/l
Suspended Solids:	90 mg/l

The foregoing reflects primary treatment removals of approximately 30% for BOD_5 (Total) and 55% for suspended solids. It has also been assumed that there is sufficient land available for construction of any of the potential processes; that the land is of suitable contour to permit flow through the secondary facilities by gravity; that existing chlorine contact tanks may be incorporated into the process without reconstruction; and that all excavation may be performed by machines, with no hard or difficult rock or subsurface conditions to interfere with construction. From the standpoint of unit redundancy, in general, two final clarifiers have been provided for the 3 MGD case, four final clarifiers have been provided for the 5, 10 and 30 MGD cases; and six final clarifiers have been provided for the 50 MGD case. Operating costs have been estimated on the basis of $0.045 per kilowatt-hour (KWH) of electrical power consumed; labor costs have been estimated on the basis of $10.00 per man-hour; and polymer costs have been estimated on the basis of $1.75 per pound of dry polymer consumed. All present worth calculations have been made on the basis of a discount rate of 7-1/8% for a period of 20 years. The following sections will briefly summarize the basic design criteria used for the development of cost estimates for the various generic process analyzed.

RBC PROCESSES

RBC processes fall into the general category of attached film processes and utilize corrugated plastic media stacks mounted on shaft assemblies. RBC equipment units are available in shafts of various standard lengths and in either standard density or high density designs. Shaft rotation may be produced either by a mechanical drive assembly or an air drive assembly. With mechanical drive RBC's, each shaft has its own motor, gear box, drive chain, etc. and with air drive RBC's, each shaft is rotated by the buoyancy of air which is captured in air cups attached to the perimeter of the media. Air is supplied from a centralized blower system through a pipe and diffuser network. For ease of comparison the estimates presented herein for air drive and mechanical drive RBC processes have been based on equipment of the same manufacturer and the design parameters have been developed on the basis of the manufacturer's published design procedures.[1] Four variations are possible for the RBC processes and are included herein. They are as follows:

1. Air Drive RBC using all high density media shafts.
2. Air Drive RBC using an appropriate combination of standard and high density media shafts.
3. Mechanical Drive RBC using an appropriate combination of standard and high density media shafts.
4. Mechanical Drive RBC using all standard density media shafts.

For all RBC processes, shaft lengths of 25 feet have been assumed and an allowance for fiberglass covers has been included. The basic design loadings have been selected at levels appropriate to meet the effluent limitation of 30 mg/l BOD_5 at a wastewater temperature of 47°F. Final clarifiers have been designed on the basis of an average surface overflow rate of 800 gallons/day/square foot. The basic design parameters used in this study are as follows:

Organic Loading:	Air Drive	- 1.7 $\#BOD_5$ (soluble)/1,000 ft^2 - Day
	Mechanical Drive	- 1.6 $\#BOD_5$ (soluble)/1,000 ft^2 - Day
Hydraulic Loading:	Air Drive	- 2.4 gallons/ft^2 - Day
	Mechanical Drive	- 2.25 gallons/ft^2 - Day

No provision has been made for separate thickening facilities for the RBC processes considered herein. Allowances have been made for providing the necessary pumping equipment, piping, etc. for returning the waste sludge to the head end of the primary clarifiers for cosettling with the primary sludge. It is anticipated that this approach would yield a combined sludge stream having a concentration in the range of 4 to 6% solids by weight. This would normally be suitable for most solids handling schemes and, further, would be comparable to the blended sludge concentrations which would be expected to result from thickening of the waste activated sludges separately and then blending them with the primary sludge.

The estimated total construction costs, miscellaneous costs, annual operation and maintenance costs and total present worth for each of the RBC processes at each of the five design flows are presented in Table I. Included in the construction costs are estimates for the RBC tankage and equipment; final clarifiers; waste sludge pumping facilities; exterior and underground piping; blower and control building (as required); and miscellaneous modifications and/or re-piping necessary to enable cosettling of RBC waste sludge with primary sludge in the primary clarifiers. As can be seen from Table I, the air drive RBC processes exhibit a slightly lower present worth than mechanical drive RBC processes with the differential present worth between the various designs becoming larger at higher design flows. The present worth of air drive and mechanical drive RBC processes using combinations of standard and high density media shafts are generally within 10% of each other throughout the range of design flows studied.

AIR ACTIVATED SLUDGE PROCESSES

Air activated sludge processes are among the oldest of the processes included in this analysis and represent the most commonly used process for domestic wastewater treatment in the past. There are many variations among air activated sludge processes which differ in their removal efficiencies; required detention times; methods of feed introduction; food to microorganism ratios; recycle ratios; organic loading; solids residence times and many other parameters. As discussed earlier, the purpose of this study is not to present the absolute optimal design for any particular process but, rather, to present a typical facilities design for each process in order to establish a basis for comparison. Other designs could be developed which might yield slightly different configurations and/or construction/operating costs; however, such a detailed presentation is beyond the scope of this study. The actual design conditions and effluent limitations will usually dictate the appropriate choice of design criteria and should be carefully evaluated on a case by case basis.

Air activated sludge processes utilize a suspended biomass, rather than an attached film as used in RBC processes, to effect wastewater treatment. The principal source of the oxygen necessary for biological activity is from atmospheric air. Since air contains only about 21% oxygen, it is necessary to compress large volumes of gas consisting mostly of nitrogen and other inert materials in order to deliver the required quantity of oxygen to the process. Further, due to the reduced saturation value of oxygen dissolved in wastewater from air, these processes are inherently less efficient

TABLE I

COST-EFFECTIVENESS ANALYSIS
ROTATING BIOLOGICAL CONTACTOR PROCESSES
(ESTIMATED YEAR 1980 COST IN $1,000'S)

	AIR DRIVE RBC		MECHANICAL DRIVE RBC	
	All High Density Media	Combined Media	Combined Media	All Standard Density Media
DESIGN FLOW = 3 MGD				
Construction Cost	$ 1,460	$ 1,540	$ 1,430	$ 1,485
Contingencies, Misc. (25%)	365	385	360	370
TOTAL COST	$ 1,825	$ 1,925	$ 1,790	$ 1,855
Annual Operation & Maint.	$ 72.6	$ 72.6	$ 93.6	$ 95.4
TOTAL PRESENT WORTH	$ 2,585	$ 2,685	$ 2,770	$ 2,855
DESIGN FLOW = 5 MGD				
Construction Cost	$ 2,210	$ 2,295	$ 2,410	$ 2,410
Contingencies, Misc. (25%)	550	575	600	600
TOTAL COST	$ 2,760	$ 2,870	$ 3,010	$ 3,010
Annual Operation & Maint.	$ 108.8	$ 110.4	$ 138.2	$ 138.2
TOTAL PRESENT WORTH	$ 3,900	$ 4,030	$ 4,460	$ 4,460
DESIGN FLOW = 10 MGD				
Construction Cost	$ 3,490	$ 3,800	$ 3,800	$ 3,965
Contingencies, Misc. (25%)	870	950	950	990
TOTAL COST	$ 4,360	$ 4,750	$ 4,750	$ 4,955
Annual Operation & Maint.	$ 162.2	$ 166.7	$ 190.9	$ 195.4
TOTAL PRESENT WORTH	$ 6,060	$ 6,500	$ 6,755	$ 7,005
DESIGN FLOW = 30 MGD				
Construction Cost	$ 8,330	$ 8,920	$ 9,710	$10,165
Contingencies, Misc. (25%)	2,080	2,230	2,430	2,540
TOTAL COST	$10,410	$11,150	$12,140	$12,705
Annual Operation & Maint.	$ 333.1	$ 334.4	$ 390.3	$ 403.8
TOTAL PRESENT WORTH	$13,905	$14,660	$16,235	$16,940
DESIGN FLOW = 50 MGD				
Construction Cost	$13,370	$14,525	$15,780	$16,445
Contingencies, Misc. (25%)	3,340	3,630	3,945	4,110
TOTAL COST	$16,710	$18,155	$19,725	$20,555
Annual Operation & Maint.	$ 513.0	$ 522.0	$ 611.6	$ 634.1
TOTAL PRESENT WORTH	$22,090	$23,630	$26,140	$27,210

in terms of oxygen transfer than are processes utilizing relatively pure oxygen, as discussed later. A variety of dissolution equipment is available for use with air activated sludge systems. Three variations are included in this study which represent a reasonable cross-section of the available equipment classifications. They are as follows:

1. Fine Bubble Diffused Aeration - This alternative uses porous ceramic plates located on an air piping distribution grid at the bottom of the tank which introduce a uniform system of fine bubbles throughout the biological contact tanks. The assumed oxygen transfer efficiency for this equipment used in this study is 26.5%.

2. Coarse Bubble Diffused Aeration - This alternative uses open-type diffusers located along a pipe header at the bottom of the tank which introduce coarse bubbles throughout the biological contact tanks. The assumed oxygen transfer efficiency for this equipment used in this study is 9.5%.

3. Draft-Tube Mechanical Aeration - This alternative uses a combined system of blowers, air spargers and mechanical mixers to effect oxygen transfer. The mechanical mixer consists of a draft tube with an internal impeller which forces wastewater downward from the surface of the tank and accelerates it past an air sparge ring where relatively coarse bubbles are broken up and entrained in the downward flow. The assumed oxygen transfer efficiency for this equipment used in this study is 50%.

For air activated sludge processes, the final clarifiers have been designed on the basis of an average surface overflow rate of 600 gallons/day/square foot for the 3, 5 and 10 MGD cases and 800 gallons/day/square foot for the 30 and 50 MGD cases. This is due to the fact that the anticipated peak:average flow ratios are, for the lower design flow cases, such that the performance of the final clarifier in terms of suspended solids removals may be adversely affected to the extent that the higher surface overflow rate may be inadequate for clarification during these peak flow periods. The remaining basic design parameters used in this analysis for air activated sludge processes are as follows:[2]

	Diffused Aeration	Mechanical Aeration
Food: Microorganism Ratio (#BOD$_5$/#MLVSS-Day):	0.27	0.35
Organic Loading (#BOD$_5$/1,000 ft^3-Day):	42	42
Hydraulic Detention Time (Hours):	5.0	5.0
MLSS (mg/l):	3100	2400
MLVSS (mg/l):	2480	1920
Average Recycle Ratio:	60%	60%

Thickening of waste activated sludge is included in the cost estimates for air activated sludge processes. Dissolved air flotation units have been provided and sized on the basis of 2.25 #/hour/square foot of flotation area.[3] One complete standby unit has been incorporated into the design to provide sufficient reliability; for the accommodation of peak loadings; and to enable the plant personnel to reduce the operating time required for thickening by using all units as may be desired. Polymer conditioning equipment has been included and the estimated consumption of polymer has been assumed to average 4 pounds of dry polymer per ton of dry solids processed.

The estimated total construction costs, miscellaneous costs, annual operation and maintenance costs and total present worth for each of the air activated sludge processes at each of the five design flows are presented in Table II. Included in the construction costs are estimates for the aeration tankage and equipment; final clarifiers; return and waste sludge pumping facilities; exterior and underground piping; blower and control building; and dissolved air flotation facilities. As can be seen from Table II, the fine bubble diffused aeration and draft-tube mechanical aeration designs exhibit similar values for present worth, while the coarse bubble diffused aeration facilities have a slightly higher present worth value, ranging from 3 to 10% higher with the differential increasing at higher design flows. This is due to the significantly lower oxygen transfer efficiency inherent in the coarse bubble diffused air systems which requires more power for the process blowers.

PURE OXYGEN ACTIVATED SLUDGE PROCESSES

In pure oxygen activated sludge processes, as with air activated sludge processes, the treatment of wastewater is effected by bringing it into intimate contact with a suspended mass of biological microorganisms within a tank designed to provide sufficient detention time to assure that the desired degree of treatment will be attained. These treatment processes utilize pure oxygen rather than air as the source of oxygen for use in biological metabolism and may have certain cost advantages over air activated sludge processes, particularly in large applications. Due to the fact that essentially pure oxygen (90-95%) is used, the natural driving force available for dissolution of oxygen into the biological suspension is much greater than would be the case using air. Using pure oxygen, the saturation value of oxygen dissolved in wastewater is approximately 4.4. times greater than with air. This further enables the active biomass to be maintained at higher levels than is feasible with conventional air systems; permits higher dissolved oxygen levels to be maintained in the mixed liquor; and permits the reduction in the volume of the biological contact tanks, thereby reducing the construction costs of these process units. From an operating cost standpoint, this results in a reduction of the amount of electrical energy required for the dissolution system in comparison with conventional air systems.

The principal drawback to pure oxygen activated sludge processes is due to the fact that the air supply used in conventional air systems is free and the high purity oxygen which is required in oxygen systems, unfortunately,

TABLE II

COST-EFFECTIVENESS ANALYSIS
AIR ACTIVATED SLUDGE PROCESSES
(ESTIMATED YEAR 1980 COST IN $1,000'S)

	Fine Bubble Diffused Aeration	Coarse Bubble Diffused Aeration	Draft-Tube Mechanical Aeration
DESIGN FLOW = 3 MGD			
Construction Cost	$ 2,085	$ 2,050	$ 1,995
Contingencies, Engr., Misc. (25%)	520	510	500
TOTAL COST	$ 2,605	$ 2,560	$ 2,495
Annual Operation & Maintenance	$ 141.4	$ 150.0	$ 138.0
TOTAL PRESENT WORTH	$ 4,090	$ 4,135	$ 3,945
DESIGN FLOW = 5 MGD			
Construction Cost	$ 2,975	$ 2,930	$ 2,870
Contingencies, Engr., Misc. (25%)	745	730	720
TOTAL COST	$ 3,720	$ 3,660	$ 3,590
Annual Operation & Maintenance	$ 190.6	$ 212.6	$ 190.1
TOTAL PRESENT WORTH	$ 5,720	$ 5,890	$ 5,585
DESIGN FLOW = 10 MGD			
Construction Cost	$ 4,770	$ 4,705	$ 4,640
Contingencies, Engr., Misc. (25%)	1,190	1,175	1,160
TOTAL COST	$ 5,960	$ 5,880	$ 5,800
Annual Operation & Maintenance	$ 299.8	$ 345.1	$ 300.1
TOTAL PRESENT WORTH	$ 9,105	$ 9,500	$ 8,950
DESIGN FLOW = 30 MGD			
Construction Cost	$ 8,530	$ 8,405	$ 8,380
Contingencies, Engr., Misc. (25%)	2,130	2,100	2,095
TOTAL COST	$10,660	$10,505	$10,475
Annual Operation & Maintenance	$ 601.6	$ 747.0	$ 608.6
TOTAL PRESENT WORTH	$16,970	$18,340	$16,860
DESIGN FLOW = 50 MGD			
Construction Cost	$13,040	$12,850	$12,860
Contingencies, Engr., Misc. (25%)	3,260	3,210	3,215
TOTAL COST	$16,300	$16,060	$16,075
Annual Operation & Maintenance	$ 912.0	$ 1,160.0	$ 924.6
TOTAL PRESENT WORTH	$25,870	$28,230	$25,775

is not. Generally, the only available means for obtaining oxygen is by providing on-site generation facilities. These facilities require a substantial initial capital investment and use large amounts of electrical energy for the production of oxygen. Therefore, a portion of the savings in electrical energy for dissolution (in some cases all of the savings) is eliminated through the addition of the oxygen generation facilities. In large plants, however, an operating savings in electrical energy consumption will generally result using pure oxygen systems. This savings often offsets the additional capital costs over the planning period enough to make pure oxygen activated sludge processes attractive.

Two principal variations are available with pure oxygen activated sludge processes. One uses covered tanks with an enriched atmosphere in the space above the liquid and one uses open tank construction. The covered tank alternative is staged into a series of three compartments and transfers 90% of the oxygen supplied to the tank through the three stages. The open tank alternative provides 90% oxygen transfer within the liquid depth by introducing the oxygen in extremely fine bubbles near the bottom of the tank. Further, two types of oxygen generation units are available: one a cryogenic generator, the other a pressure swing adsorption (PSA) generator. In general, a cryogenic generation system costs more initially but requires less electrical energy than does a PSA generation system. Selection between these two types of systems within the range of design flows included is not always obvious and, therefore, both will be presented for illustration purposes. The four variations included in this study are as follows:

1. Open Tank construction with a cryogenic oxygen generation system.
2. Covered Tank construction with a cryogenic oxygen generation system.
3. Covered Tank construction with a PSA oxygen generation system.
4. Open Tank construction with a PSA oxygen generation system.

For pure oxygen activated sludge processes, the final clarifiers have been designed on the basis of an average overflow rate of 600 gallons/day/square foot for all cases included in this study. Although with air activated sludge processes it was deemed appropriate to increase the design overflow rate at the higher design flows, it is felt that solids flux considerations could be controlling in the pure oxygen activated sludge processes and that the more conservative design is indicated within the range included in this study. The remaining basic design parameters used in this analysis for pure oxygen activated sludge processes are as follows:[4,5]

	Open Tanks	Covered Tanks
Food:Microorganism Ratio (#BOD$_5$/#MLVSS-Day):	0.70	0.75
Organic Loading (#BOD$_5$/1,000 ft^3-Day):	155	175
Hydraulic Detention Time (Hours):	1.35	1.2

	Open Tanks	Covered Tanks
MLSS (mg/l):	4,375	4,575
MLVSS (mg/l):	3,500	3,660
Average Recycle Ratio:	40%	40%

Thickening of waste activated sludge is included in the cost estimates for pure oxygen activated sludge processes. Dissolved air flotation units have been provided and sized on the basis of 3.0 #/hour/square foot of flotation area.[3] As with air activated sludge processes, one complete standby unit has been incorporated into the design. Polymer conditioning has been included and the estimated consumption of polymer has been assumed to average 4 pounds of dry polymer per ton of dry solids processed. The average underflow concentration from the clarifiers has been assumed to be 1.75% solids by weight. In the larger cases studied, centrifugation also was considered for thickening. In the 30 MGD case, a supplementary cost-effectiveness analysis indicated that the dissolved air flotation and centrifuge alternatives were comparable in terms of present worth and in the 50 MGD case, the centrifuge exhibited a slight present worth advantage; however, since centrifuge data was based upon certain assumed polymer dosages and loading rates which would only be verifiable through pilot testing, it was believed more realistic to presume the use of dissolved air flotation for thickening in all cases studied.

The estimated total construction costs, miscellaneous costs, operation and maintenance costs and total present worth for each of the pure oxygen activated sludge processes at each of the five design flows are presented in Table III. Included in the construction costs are estimates for the oxygenation tankage and equipment; final clarifiers; return and waste sludge pumping facilities; exterior and underground piping; oxygen control building; oxygen generation and storage facilities; and dissolved air flotation facilities. As can be seen from Table III, generally all of the pure oxygen activated sludge facilities exhibit almost the same present worth for any given design flow, within very close limits. There is a slight consistent advantage favoring the PSA oxygen generator alternatives, although it is not significant. In general, the selection of oxygen generator type would be more reasonably made on the basis of system needs, such as the turndown requirements. Cryogenic generators are usually limited in turndown to about 60 to 70% of rated capacity, while PSA generators may be more or less turned down almost to zero output.

COMPARATIVE ANALYSES

In conducting cost-effectiveness analyses, the annual operation and maintenance costs play an important role in determing the selected alternative. These costs are converted into equivalent present worth by multiplying by an appropriate factor (10.49187 in this study). The single most significant annual operating cost, particularly for the larger design flows, is the cost of power. Table IV presents the total annual power requirements for each of the processes and design flows considered in this study. In general, the RBC processes have a significant advantage in terms of annual

TABLE III

COST-EFFECTIVENESS ANALYSIS
OXYGEN ACTIVATED SLUDGE PROCESSES
(ESTIMATED YEAR 1980 COST IN $1,000'S)

	CRYOGENIC GENERATOR		PSA GENERATOR	
	Open Tanks	Covered Tanks	Covered Tanks	Open Tanks
DESIGN FLOW = 3 MGD				
Construction Cost	$ 2,810	$ 2,705	$ 2,575	$ 2,595
Contingencies, Misc. (25%)	700	675	645	650
TOTAL COST	$ 3,510	$ 3,380	$ 3,220	$ 3,245
Annual Operation & Maintenance	$ 151.9	$ 149.9	$ 152.4	$ 152.8
TOTAL PRESENT WORTH	$ 5,105	$ 4,955	$ 4,820	$ 4,850
DESIGN FLOW = 5 MGD				
Construction Cost	$ 3,695	$ 3,495	$ 3,355	$ 3,440
Contingencies, Misc. (25%)	925	875	840	860
TOTAL COST	$ 4,620	$ 4,370	$ 4,195	$ 4,300
Annual Operation & Maintenance	$ 204.7	$ 201.3	$ 205.6	$ 206.2
TOTAL PRESENT WORTH	$ 6,770	$ 6,480	$ 6,350	$ 6,465
DESIGN FLOW = 10 MGD				
Construction Cost	$ 5,385	$ 5,330	$ 5,160	$ 5,090
Contingencies, Misc. (25%)	1,345	1,330	1,290	1,270
TOTAL COST	$ 6,730	$ 6,660	$ 6,450	$ 6,360
Annual Operation & Maintenance	$ 317.6	$ 311.6	$ 319.2	$ 320.8
TOTAL PRESENT WORTH	$10,060	$ 9,930	$ 9,800	$ 9,725
DESIGN FLOW = 30 MGD				
Construction Cost	$ 9,475	$ 9,600	$ 9,340	$ 9,035
Contingencies, Misc. (25%)	2,370	2,400	2,335	2,260
TOTAL COST	$11,845	$12,000	$11,675	$11,295
Annual Operation & Maintenance	$ 614.8	$ 598.6	$ 619.1	$ 623.1
TOTAL PRESENT WORTH	$18,295	$18,280	$18,170	$17,830
DESIGN FLOW = 50 MGD				
Construction Cost	$14,160	$13,970	$13,610	$13,570
Contingencies, Misc. (25%)	3,540	3,490	3,400	3,390
TOTAL COST	$17,700	$17,460	$17,010	$16,960
Annual Operation & Maintenance	$ 916.7	$ 890.8	$ 923.4	$ 929.8
TOTAL PRESENT WORTH	$27,320	$26,805	$26,700	$26,715

TABLE IV

COMPARATIVE POWER REQUIREMENTS
(ESTIMATED POWER - 1,000 KWH/YEAR)

PROCESS	3 MGD	5 MGD	10 MGD	30 MGD	50 MGD
ROTATING BIOLOGICAL CONTACTORS					
1. Air Drive - All High Density Media	245	420	755	2,055	3,445
2. Air Drive - Combined Media	245	455	855	2,085	3,645
3. Mechanical Drive - Combined Media	250	500	820	2,420	4,045
4. Mechanical Drive - All Standard Media	290	500	920	2,720	4,545
AIR ACTIVATED SLUDGE					
1. Find Bubble Diffused Aeration	550	855	1,620	4,755	7,860
2. Coarse Bubble Diffused Aeration	850	1,455	2,850	8,320	13,925
3. Draft-Tube Mechanical Aeration	585	955	1,850	5,245	8,695
PURE OXYGEN ACTIVATED SLUDGE					
1. Open Tanks (Cryogenic Oxygen)	560	945	1,795	4,715	7,630
2. Covered Tanks (Cryogenic Oxygen)	515	870	1,660	4,355	7,055
3. Covered Tanks (PSA Oxygen)	570	965	1,830	4,810	7,780
4. Open Tanks (PSA Oxygen)	580	980	1,865	4,900	7,920

power consumption. Throughout the range of design flows analyzed herein, the RBC processes generally require from 40 to 65% of the power required for pure oxygen activated sludge processes and from 23 to 60% of the power required for air activated sludge processes. Further, air drive RBC processes generally consume somewhat less power than do mechanical drive RBC processes.

Table V presents the total present worth for each of the processes and design flows considered in this study. For the 3, 5 and 10 MGD cases, all RBC processes exhibit a significant cost advantage over the air and pure oxygen activated sludge processes. At 30 and 50 MGD, the air drive RBC processes have a significant cost advantage over mechanical drive RBC processes and the air and pure oxygen activated sludge processes, all of which are closely grouped, with the exception of the coarse bubble diffused aeration version of the air activated sludge process.

Following the preparation of the cost-effectiveness analysis, it may be necessary or desirable to examine the subjective considerations for those processes which are grouped within 5 or 10% of the present worth of the alternative having the lowest present worth value prior to selecting the final plan. Based on the information presented herein, for the 3, 5 and 10 MGD cases, this would mean the comparison of air drive and mechanical drive RBC processes and for the 30 and 50 MGD cases, this would mean the comparison of essentially all of the alternatives with the exception, perhaps, of the coarse bubble diffused aeration version of the air activated sludge process. In terms of operational simplicity; power requirements; waste sludge thickening; and labor costs; all of the RBC processes have, in the author's opinion, a definite subjective advantage. The various activated sludge alternatives may have somewhat of a subjective advantage in terms of flexibility; reliability; and shock load assimilation capacity; however, not a significant enough advantage to warrant the skewing of the analyses. Land area requirements are comparable for RBC and air activated sludge processes; however, pure oxygen activated sludge processes have the advantage there. In terms of comparing the air drive and mechanical drive RBC processes, the air drive RBC processes have, in the author's opinion, the advantage in terms of maintenance, power requirements and turndown capability. The foregoing illustrates some of the various subjective considerations which may be considered following the completion of the cost-effectiveness analysis step. None of these considerations are sufficiently important to reverse the results of the analysis and, therefore, it is recommended that they be addressed briefly, for completeness, and then the final selection should be made based on the results of the cost-effectiveness analysis.

DISCUSSION AND CONCLUSIONS

Recognizing that the estimates contained herein, although rigorously developed, are budgetary in nature and not site specific for any particular set of circumstances, the authors acknowledge that variations are to be expected from the costs presented in this study. There are many assumptions built into these cost estimates and they may not necessarily suit the preferences of other analysts or be appropriate for a given situation. The purpose of this study has been to investigate, over a significant range of

TABLE V

COMPARATIVE COST-EFFECTIVENESS
(ESTIMATED TOTAL PRESENT WORTH IN $1,000'S)

PROCESS	3 MGD	5 MGD	10 MGD	30 MGD	50 MGD
ROTATING BIOLOGICAL CONTACTORS					
1. Air Drive - All High Density Media	$2,585	$3,900	$6,060	$13,905	$22,090
2. Air Drive - Combined Media	$2,685	$4,030	$6,500	$14,660	$23,630
3. Mechanical Drive - Combined Media	$2,770	$4,460	$6,755	$16,235	$26,140
4. Mechanical Drive - All Standard Media	$2,855	$4,460	$7,005	$16,940	$27,210
AIR ACTIVATED SLUDGE					
1. Fine Bubble Diffused Aeration	$4,090	$5,720	$9,105	$16,970	$25,870
2. Coarse Bubble Diffused Aeration	$4,135	$5,890	$9,500	$18,340	$28,230
3. Draft-Tube Mechanical Aeration	$3,945	$5,585	$8,950	$16,860	$25,775
PURE OXYGEN ACTIVATED SLUDGE					
1. Open Tanks (Cryogenic Oxygen)	$5,105	$6,770	$10,060	$18,295	$27,320
2. Covered Tanks (Cryogenic Oxygen)	$4,955	$6,480	$ 9,930	$18,280	$26,805
3. Covered Tanks (PSA Oxygen)	$4,820	$6,350	$ 9,800	$18,170	$26,700
4. Open Tanks (PSA Oxygen)	$4,850	$6,485	$ 9,725	$17,830	$26,715

design flows, the relative cost-effectiveness of RBC processes in relation to activated sludge processes for a typical carbon oxidation application. The results of this study indicate that RBC processes are, indeed, cost-competitive with activated sludge processes throughout the range of design flows investigated. Therefore, the RBC processes should be given careful consideration in any similar application.

Further investigation is warranted in several areas with regard to RBC processes and cost-effectiveness in general which either surfaced during the preparation of this study or were observed by the authors during its preparation. First of all, the range of design flows could be extended upwards, perhaps to 100 MGD, since no distinct breakpoint was observed beyond which RBC processes were no longer preferred on the basis of their present worth values. Secondly, the use of inclined plate-settler types of clarifiers may be warranted and could greatly reduce the costs for RBC processes. Use of the plate settlers would result in the additional benefit of reducing land area requirements for the RBC processes to levels comparable with pure oxygen activated sludge processes. The need for waste sludge thickening may require additional study to determine whether or not it is required as an additional cost item for RBC processes or as a deduction from activated sludge processes. Further, as more data is generated from centrifuge manufacturers and/or operating installations, the use of dissolved air flotation exclusively for thickening as presented in this study may require reexamination.

In conclusion, this study indicates that RBC processes are viable alternatives for carbon oxidation; are cost-competitive and cost-effective throughout the range from 3 to 50 MGD and should be carried through the detailed analysis performed as a part of future facilities planning work for proposed BOD_5 removal projects.

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WORKSHOP ON RBC RESEARCH NEEDS *

Chairman: Dr. J. A. Borchardt
Assistants: Dr. Y.C. Wu, Dr. Ed. D. Smith, Dr. R. D. Miller,
Mr. E. J. Opatken, Dr. W. A. Sack, Dr. C.P.L. Grady, Jr.,
Dr. S. K. Banerji, Dr. D. F. Kincannon, Dr. A. A. Friedman,
Dr. W. G. Characklis, Dr. F. M. Saunders, Dr. J. C. Haung,
Dr. P.C. Poon, Mr. R. L. Antonie, Mr. G. R. Fisette,
Mr. G. E. Flann, Mr. J. T. Madden, Mr. B. Joost, Mr. M. Creston,
Dr. C. G. Steiner, Prof. W. W. Eckenfelder, Mr. J. F. Lagnese,
Dr. E. J. LaMotta

DR. BORCHARDT: We are very pleased and grateful to see so many of you come and I know there's a lot of competing interest for your time. We hope that this will be a productive meeting. I thought perhaps that we would organize this in such a fashion that we would address five separate topics and for that purpose we have picked five subchairmen to each discuss one of these topics for a few minutes. And then we are going to, for that particular subject, call for discussions, questions, statements from the floor for another 10 to 15 minutes. That allows about 20 to 25 minutes for each subject, hopefully that five of them you see we should finish in an hour and a half or something of that sort. So that is the plan for the evening. Now the subjects and subchairmen that we are going to use, we have three of them here on the podium and I'm expecting the other two any minute. The first subject is scale factors from pilot to prototype and Professor Grady from Purdue University is going to be the subchairman in charge of that particular subject. Professor Grady is here on my left on the end at the present. The next subject we are going to talk about; kinetics, internal and external diffusion problems. Professor LaMotta is going to take that subject. Then the third group will be RPM turbulence surface effects--that sort of thing, and Professor Friedman has agreed to chair that; he's not here yet as I see it, and then operating problems. Mr. Legnese of Duncan-Legnese Associates is going to take the operating problems. And then we have research support and I know you are all interested in research support and Ed Opatken has agreed that he will discuss this and answer all questions. So you see, you have a full evening ahead of you and I hope that you are not thinking about other problems, that you're concentrating on this research. If you were here yesterday you heard about Murphy's Law and O'Toole's corollary and a few other things. I suppose you all realize that Mrs. Murphy has a law as well. Mrs. Murphy's law says in effect that if anything can go wrong it will go wrong and when Dad is out of town, so you can expect the plumbing is broken or the fuses are blown or something at home, but just relax until the meeting is over and then worry about those problems. Now we still don't have our other two subchairmen but I'm not going to worry about that for I know we have enough experts in the group here to handle them. So our first subchairman tonight will be Dr. Grady from Purdue University. He is going to talk about scale-up factors from model to prototype.

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*The discussions from this workshop were typed from a tape of the workshop. Consequently, problems with content and clarity may be apparent in certain sections.

DR. GRADY: I am going to talk very briefly hopefully because I feel on scale factors there is a very strong interaction. I am talking about scale factors. I am talking about each of the second and third topics today, because really scale factors or scale-up is basically a problem of identifying the fundamentals within the system so that we can then know how those fundamentals are affected by system signs. You really can't talk about scale up without talking about turbulence diffusion and all of these other problems. So I guess what I would like to do for a moment is to perhaps reiterate some of the things that I said this afternoon, some of you I know were not here. The course of that discussion was that by using relatively fundamental mathematical models of the type that Jim Mueller uses and other people are working on within this field, you can take into account scale factors. Obviously all of you have had fluid mechanics background somewhere along your way and you know that scale factors are handled in a number of ways in a fluid system. There is really no reason why we cannot handle scale factors in the same way within our system even though they are going to be more complex because they involve more factors. But since my point of view and I guess I am trying to give you something to start, to get the discussion going, because certainly I do not have any answers on the thing. But from my point of view the best way to approach scale-up is first to develop fundamental mathematical models and then to use those fundamental mathematical models to study the effect of scale for a large number of different situations. Only through doing that can one begin to see generality which can then be simplified so that those rather complex mathematical models can be reduced to simpler forms which contain the most important factors. Then once that has been done, we can then go about rationally designing prototype studies so that those prototype studies can then be scaled up. So one did ask the question what exactly do we mean by scale-up? Are we strictly talking about size of the system or are we also talking about process engineering of the systems. And I would hope as you consider the question of scale-up, that you will also consider the question of process engineering because I heard a lot of people asking questions this afternoon about such factors as what are optimal feed policies? What are optimal size reactors for different stages within the system? These are questions that can be answered for any given system again once you have a relatively fundamental mathematical model, because then you can play with that system and can come up with the either feed pattern, with the loading pattern, with the reactor configurations that would be most cost effective for any given set of kinetic parameters for that system. In addition, one of the areas of research in the field of chemical engineering right now which is very important as far as we are concerned is how to take into account uncertainties during design and again once you have a fundamental mathematical model for the process then you can begin to apply some of the new research techniques to consider the fact that all of our parameters which go into these models will have uncertainty associated with them. In addition, we all face uncertainty when it comes to loading on our plant, what will happen in the future on those plants. So there is a growing body of knowledge now as to how to handle those things in the design process. One will hope that this also will become a research need or a research application within our field as well. So as I look at scale-up then, I see scale-up in a broader sense and research needs for scale-up in a broader sense than just taking a prototype unit and figuring how to increase its size. I would hope that we would be doing research that will tell us more about mass transfer coefficients, how they are influenced by rotational speed, by disk diameter. I would hope that we would be doing research that helps us know more about oxygen transfer within the system because once we know that, then we can look at the benefits that would come from air drive or pure oxygen systems as Jim Mueller talked about this afternoon. I would hope that we would continue to do research on the effect of rotational speed and disk configuration on the fitness of the liquid layer that is carried up into the aerated sector. I would hope that we could continue to do research on the effect of disk configuration on the distribution of that layer since it is going to be moving by gravitational forces. I would hope that we would continue to do research on the biomass thickness within the system as a function of rotational speed and loading upon that system. I would imagine that the other speakers will discuss that problem as well. So I feel at a point in time where we have an opportunity to develop an understanding of these systems that would help us make extrapolations which have not been possible in our field in general in the past. And with that I guess I will ask you for your comments because certainly you have a lot more experience with scale up than I do. So with that I will turn it over back to Jack and let him fill your comments and your questions.

DR. BORCHARDT: We have five microphones available, one in the back, one here, one up front here and one over here. Please avail yourself with these microphones, give your name and state your question if you have a question, or statement. If you have had a problem or whatever it is, we would like to know it. This is your chance to express yourself. Is there anyone who would like to make a comment relative to scaling of pilot to prototype?

DR. SMITH: There is one obvious scale-up factor that, at least I think it's very obvious, but I hear very rarely anyone ever mentioning it and it is the fact that the matrix design for instance you can design, say you need 24 units. Are you going to put it in a two by twelve matrix or three by eight matrix, four by six matrix, six by four matrix, etc., one by 24, and it is very important based on my experience and some other people here when we talked about it. This is very important because if you choose the wrong matrix, often times you overload your first units and they will go anaerobic and therefore you will lose your nitrification capacity, etc. I think this is a very important scale-up problem and should be addressed. Does anybody else have any comment on that?

DR. GRADY: Well, that's the thing I was talking about Ed, when I was talking about reactor engineering within our system, is because once you have a decent mathematical model then that's what it allows you to do and you can look at all of those combinations with a very small capital outlay as far as operating costs on the computer is concerned. And then it just depends upon how much faith you have in the model.

MR. HENNESSY: One of the things that I have noticed today is that people have not even determined what the real parameters are that determine real operation of the bio-surf and even at that they have not even agreed what the answers are once they come up with the parameters. What I would like to know is do you really feel that at this time you can walk into a treatment, a sewage plant that has to be upgraded and merely say this is a cost effective system and we can put it into operation and it will do this for you. You know, can you really guarantee that you are spending the people's money properly?

MR. WONG: I am addressing to the manufacturers. We have many, many manufacturing plants and we are going to use the biological discs data and from what I have heard the last two days I am not too sure where I stand from the standpoint that we have facility to be constructed. We know we will have certain waste streams to be treated and we run a pilot study with 2-ft. discs and I wondered...and we of course, we run it at a very specific configuration that we take and specific conditions that we take, and I am just wondering whether our tentative data package is sufficiently comprehensive to be capable of going out on open bids to the manufacturing communities. It seems to me from the discussion and from the presentation that it almost necessitates that we have to tailor our package to the pilot work that we have done and know what should be the configuration of the discs that we have used and to the type of flow rate and drawing rate, etc.. And we just come out with the generalities. I wonder whether such a package would be sufficient for competitive bids. I don't know how the manufacturers that are represented here view these conditions.

MR. HYNEK: It's a very good question and I would refer back to Al Friedman's presentation this afternoon that he has raised very logical questions, what is the work of a pilot plant, and we've discovered the same thing. A small two-foot diameter pilot plant is an excellent tool for determining what type of biology we will get, what type of removal you will get, but aside from educating the customer as to the process and I could say an advantage, simplicity of startup and quick degree of treatment. It is a learning tool particularly in the industrial sector. The customer does not realize the impact of his manufacturing process on waste treatment, and in the process of a pilot study he learns very quickly the economic impact of what he has been doing and it leads to modification and improvement of his product process. We relied very heavily on the small pilot plant because it was an effective tool to get out in the marketplace, get someone to understand the process. As soon as we could we switched to full scale data and basically our approach today is, and Jim Mueller mentioned that Autotrol funded mass modeling work, and I've been involved in that. I think this is a great step forward. Basically we use the small models today to get a feeling for it from our point of view because we have a large experience in data bank. We know what the small ones can do. We have learned from full scale installations what the big ones can do. When push comes to shove, you're coming to us for an appraisal of what the full scale plant has to be. We rely on the mass model information we have learned because it is realistic. We rely on our full scale data bank, it is realistic, and you are going to have to stop looking at ten-pound removals. You are going to have to look at the realistic world. I do not know if I have answered the question but I tried damn hard.

The use of mathematical models is a separate goodie. That work separately from actual pilot plant data. So I'm an advocate of pilot data but why emphasize the limitations of scale-up for turbulence and oxygen and that is my thought.

MR. FRIEDMAN: There was some good work done in Canada a couple of years ago on municipal waste waters where they scaled from...so I remember from half meter to about six-foot. The scale-up factor required an additional 16% area. The question then was what happens when you go prototype and I kicked this around with the primary author several times because 80% of the times, we were both scared stiff point blankly, because we do not know how to scale-up. I think both of us would argue that somewhere between 25 and 30% of linear scale factor out of ignorance.

DR. HOVEY: Personally addressing the fellow from the army's problems, you've got what...10, 15 different configurations of pilot plants and media. You got ten I don't know how many manufacturers sending things out to bid and your starting point, your pilot operation is manufacturer A, manufacturer C, a manufacturer D pilot plant. Perhaps what we need is a standard configuration from which we start and we have done it for say seven tests and all kinds of ASTM standards you know, and this is the apparatus and maybe what we need is a standard pilot plant from which we would develop our pilot data and then it becomes the manufacturer's responsibility. You know you must match this pilot plant data with his full scale materials that you are going to bid on the job with.

DR. KETCHUM: We are working on the project with RBC and we have done the pilot scale work and they are going to do the design criteria, and the big question is okay, we have the pilot scale data that contains up to a thousand mg/l of formaldehyde, has a very selected culture degrading it, and we have all this nice data. What do we do with it? For fun, you are the experts. How do we scale-up? What kind of a biodisc do we use? What is it a dissolved oxygen limited system. We know that. What kind of disc configuration is the best for dissolved oxygen limited system? What kind of loading rates are best? It's a very easy degradable type of material. It goes very rapidly. We say 70, 80 percent degradation in our flash chamber. What do you do from here? Can you go to a 50% disc space in the first chamber and be able to make it run or is it going to do with dissolved oxygen limited. What do you do with it?

DR. FRIEDMAN: I really do not have an answer for you and also I really do not believe we have an honest set of scale-up factors at the present time that we can apply.

DR. KETCHUM: We have gone out to four or five manufacturers and we have come back with different configurations from each one. I mean vastly different--nothing even resembles each other. So putting it all together, we kind of came up with a configuration of our own. And another question I would like to ask is okay, if you do a system that has 50 to 60% of its area in the first disc, we know that we can get 80 to 90% BOD reduction in that area saying that we are not dissolved oxygen limited. The pilot scale study shows that the third and fourth discs which are running in at 2-1-1 configuration had very little growth on it. The third disc removed some BOD and the fourth just went out for the ride. How can we in a full scale unit design so that we can get better loading of those discs and higher BOD reductions so we can get lower effluent BOD valve.

DR. O'SHAUGHNESSY: You started this mess, I'll let you finish it up. In the beginning when we were talking about scale-up factors and we were going round and round, and there has been one pertinent comment tonight on scale of factors, all of us are running around doing our own little thing with this unit and that unit and they are not the same, and I think if we would go into a laboratory and take any wastewater and put it in a settling column test with a six inch diameter, and go to a two and half foot diameter it would come out with different settling curves. And what do we use for scale of factor on the domestic wastewater? Well, we go and find out. Why in the heck don't we start with some standard pilot units on domestic wastewater so we can get some information to look at a certain geometry on a fullscale plant and find out what the heck scale-up factors are, instead of beating each other over the head. Forget industrial waste, we will never get that answered.

MR. WONG: As a customer I wish that various manufacturers will form a league or an association or society to come up with a typical pilot model from the standpoint that if the customer wants to have open bid on the job and if the pilot model is followed, these manufacturers will be able to bid on the job without any reluctance. If they can come up with that type of agreement among themselves, it would help the customers a lot.

DR. BORCHARDT: Now, we are going to pass on to the next subject which perhaps will be a little bit easier for us to accept, and this has to do with kinetic factors internal and external diffusion, and the discussion will be led by Dr. LaMotta, University of Massachusetts.

DR. LaMOTTA: What I would like to do is summarize the most important concepts relating to external diffusion and internal diffusion and both needs for the mathematical modelling. From my viewpoint, first of all the mass transfer can be studied in two general cases, the disc has surfaces exposed to the air which means we are getting three phases; the gas stage, liquid stage, and the biofilm, and we have another case when the disc is emerged in the trough and we only have two phases; the liquid film and the biofilm. Then if we consider first the most difficult case, with the disc in the air, we have three phases we have to consider; the transfer of oxygen from the gas phase to the liquid film, then the transfer of oxygen from the liquid film to the biological film. So we have two types of mass transfer processing, and none of them have been addressed so far in the literature. Most of the models and most of the studies have concentrated on the transfer of oxygen from the gas to the bulk solution in the trough. I do not remember seeing any paper in which we have discussed what happens with the dissolved oxygen, for instance, on the liquid film in the air. What is the effect of biological utilization of dissolved oxygen on the liquid film in the air? Well, some people may say if the rotational speed is slow then you get dissolved oxygen saturation in the liquid film. Well, if that is the case, one may think the rotational speed definitely has an effect. However, that problem has not been addressed yet. With respect to the transfer of the dissolved organic from the bulk of the liquid, when I say bulk of the liquid I mean in the liquid film, to the biological film, again we ignore exactly what is going on.

MR. WONG: I do not think the question is really answered from the standpoint of I really still do not know what to do? That this really means that, from what I heard, is a pilot is only to build the confidence of their customer from that standpoint, I really do not need the pilot. If I do not need a pilot and of course we have to prepare a tentative package to go on bid, what do manufacturers need in the tentative package to be able for them to provide the service?

DR. BORCHARDT: Well I think you must realize that Autotrol has been getting data for perhaps the past fifteen years. So now they do not need pilot plant work as much but when they first started they desperately needed pilot work.

MR. WONG: When you talk about pilot work, I think you will agree that each stream is merely different...And Autotrol I wonder if they have ever had a stream that contains explosive and repellant particles. You know it is different. I'm just wondering whether our pilot really serves the purpose. What does it intend to do?

DR. BORCHARDT: Now I think that we might discuss this all evening and not really reach a plateau of mutual agreement. There is one thing about industrial wastes such as you have mentioned, in general, they are probably sterile and you just do not turn the pilot plant on and say whatever results I get that is what the prototype will do for me. It is a question of developing a seed that seed will be happy in that environment. Now many times the environment changes suddenly maybe, but the environment does change. You have heard an awful lot about Beggiotoa. For example, I've seen plants where overnight Beggiotoa will appear and then in a similar sharp fashion go away. That means the biomass increases from 1/8 to 3/16 up to perhaps a half an inch. The mixed liquor solids jump up from 250 to 2500 purely because the substrate has changed suddenly and you cannot model that easily because it is one of those things that depends on proper seeding and so forth and especially in an industrial waste this is a difficult thing. I have run an industrial pilot for months and couldn't get the thing to start working well, even though I kept throwing seed at it. So these kinds of things do take a little bit as we call it empathy. You have to understand the system and work with it. Now I would say that is a pretty tough question to say are you spending people's money properly. So I think shouldn't really debate that. Let us say in order to meet an established standard I think that is something else again. I think we can accomplish that. I am not sure that your question meant that; however, it would be my opinion that in most cases a good consulting firm that has the tools that have been mentioned here at the meeting would certainly be capable of telling you how to meet this state of objective with this RBC process.

MR. MADDEN: It would seem to me that if we were going to start to look at scale-up, we might first address the possibility of securing data that would address scale-up as it has to be addressed, that is to say, take a plant that has similar characteristics of waste water of the various diameters or various pilot sizes and secure absolute data, hard data, from which you could then produce your mathematical models. I think to start from the mathematical model end and work toward the plant reality or the treatment facility as it will ultimately be designed, may in fact be the wrong direction. I do not think there is sufficient data available from any of the manufacturers or from any of the waste treatment plants that are now on line that can produce the quality of data required to employ mathematical models, something you can scale-up from a two foot diameter, or four, or six, or ten, and ultimately design a 24 or 60 or 80 shaft job on it. Also it should be pointed out that most of the facilities, I know speaking for our facility, most of the operational facilities that we have on line are nowhere near either design flow; they may not be near organic stream flow; they may have idiosyncracies that were not considered at the time the project was designed. So to secure that data and then force the data into a position so that you can develop a mathematical model and accept probability or a mathematical probability as the criteria for then saying it is going to work, it is going to fit, it just does not fall into...it just seems that that is not the way to go.

DR. BORCHARDT: Well I think in general we are not actually here to debate with you that particular issue if that is your conviction. In general the use of mathematical models I think is here to stay. In general what you do is use the model to indicate to you the variables that you have to define with more finesse which you then proceed to do, and then you work back and forth until you get something that seems to you to work, and in general that I think is what is happening in this field. At first we started out empirically but now we are moving more and more in the direction of getting models. I myself have used two-foot models, 18 inch, one-foot models, four-foot models and on up to the 11 and 12-foot prototypes. So that I do feel we are using models in that fashion and I feel that they are doing a great job for us in helping us. Frankly I feel that is the way we must go.

DR. MOLOF: I am an advocate of models. We had a scale-up system. We showed that RPM was the scale-up choice but you cannot do it because this couldn't take it. Now we have a way of doing it. The problem... let me just remind everyone about scale-up. The bacteria that grow on the disc of a 3-foot are the same bacteria that grow on the 12-foot disc. If we can duplicate, we don't have to duplicate the bacteria. They grow the same size. We find out on the 3-foot disc or the 2-foot disc or 1-foot disc what the properties of the system are and domestic waste changes. It has infiltration inflow, dilution stream. I've done a lot of work with different types of sewage and different types of industrial wastes. I can learn a lot by pilot, and let me explain I was in industry and I was in a big company and they were wise guys like some of you and they said let us skip the pilot plant. It went from laboratory to full scale plant. They had to write it off for several million dollars a year later. I think that construction is too expensive, that if we have any doubts, go to a small pilot plant. They are not that expensive. You learn a heck of a lot. You learn we are going to go anaerobic. You learn do we have enough oxygen. Let me just comment what we have to learn to scale. You have to learn to scale mixing. Now mixing is something that the field has been working on for 50 years. We should have that expertise in mixing so we have to study that as Dr. Grady said we have to study turbulence in these reactors; we have to study fundamentals such as oxygen transfers. Oxygen transfer is the other thing. If we can scale oxygen, we can scale turbulence. I can go from a six inch to a 12-foot and be right every time, and that is what our goal should be.

Therefore my belief is in that part of the rotating disc exposed to the air, we know nothing, absolutely nothing. And if you want to make a good progress in understanding the behavior of the biological film, and in the rotating biological contacter with the biofilm in it, we had better begin to do research in that area. Most of our research has concentrated on the second case which is the innermost portion of the disc. Still then in that case, the assumption made in most of the mathematical models has been that the mass transfer coefficient K_L and all those correlations presented in the papers. Can we directly calculate dissolved oxygen in the bulk of the liquid given that we know the geometry of the disc being considered. One of the basic assumptions in the mathematical models has been that the liquid film that we have in the air. If the oxygen transfer occurs instantaneously upon the entry in the trough that is false. As Dr. Grady said this afternoon, that is a lie. Then, in the mathematical model we make that assumption which is a lie. You see, nobody knows exactly what the effect of not making that incorrect assumption will be in the performance of the model. Therefore my suggestion would be let us make the more difficult realistic assumption that there is not any completely exchange of the liquid film in the trough. In speaking of a disc boundary layer when the disc reenters in the trough, and that disc boundary layer thickness is larger than the thickness of the liquid film of the disc in the air. So we have to study the mass transfer coefficient of the oxygen and this organic from this boundary layer, from this boundary layer to the bulk of the liquid and to the biological film. We ignore, absolutely ignore the mass transfer coefficient in this case. This is what I think the most important research needs in the future. With respect to internal diffusion, all of us knows the mathematical models have to use the coefficient of the diffusibility of the substance inside the biological film. The first big problem we have in dealing with the subject of the sewage. We asked the question, the diffusivity of what? Some of the mathematical models have used the diffusivity of glucose. But we know that is not realistic. And then the question is, should we use, if it is a compound or complex organic, like in the case of sewage where for the most part the mathematical model, modelling the diffusion of substance inside biological films. In my opinion we should try to understand first the phenomenon, using the statistics compounds that can be directly measured instead of using indirect preparation of parameters of COD, BOD, or TOC. We do not know anything about the effects of film thickness on the effective diffusivity of the substance. In most mathematical models the effective diffusivity continues to be a constant. We have known in the case of suspended growth, the effective diffusivity is the function of particle size diameter. Similarly you might expect the effective diffusivity is not a constant, it may be affected by film thickness. This knowledge has not been explored yet. If you pull everything together in a mathematical model of the process, then you have to begin considering all these complicated facts. They may lead us to a logic equation. We saw in the first equation of that happening due to the biological portion of the film exposed to the air of the liquid film exposed to the air, and we have to combine those effects with what is happening inside the trough, and we wound up with a logic differential equation. For the verification of the model, we have done by using BOD, TOC, or COD, have a parameter to measure concentration. In my viewpoint that is inappropriate, simply because BOD, TOC and COD are global parameters, operational parameters that do not let you know what is happening with each one of these specific compounds that we have indeed. So even though I advocate strongly for mathematical modelling simply to better understand what are the most important factors affecting the performance of the unit, I wouldn't recommend using directly that mathematical model as a mathematical formula and use BOD, TOC or COD as a parameter to measure concentration. With this I would like to open the discussion on several topics. I know that there are people who have not done work in this area simply because it usually involves high mathematical differential equations and most practicing engineers do not like to deal with that. However, there are some people who have done some work in this area and I would like to discuss this problem with them.

DR. GRADY: I wanted to throw out one other thing, along these same lines for those of you that are working physically in these areas to think about: that is, is it possible to come up with a chemical system, perhaps an heterogeneous catalysis system, that is much simpler from the reaction standpoint, that could serve as a prototype model that could be studied in order to get at these basic physical mass transfer parameters in a simpler manner than is required with our complex biological system.

DR. LaMOTTA: Well, in that case even the physical would make sense... there might be the criticism that we are neglecting the biological activity. Some people even believe that there is an active transport of substances endued by, say the cells which are inside of the biological form. And so we would be neglecting these effects, that is the only danger. But I believe that you are right. We may understand better the system if we work with an inert system, and we would then study only the physical parameters involved in the process.

DR. WU: In order to avoid the unnecessary argument which may occur at the second symposium, the methods or techniques employed for the measurement of the RBC system operating parameters should be defined. Otherwise, the kinetic constants obtained from different studies may be different although they are derived from the same model/equation. For this reason, the research funding agency such as EPA or NSF should look into this problem and make suggestions to those persons who are planning to conduct the RBC research in the future.

DR. CHARACKLIS: I would like to comment first just on modelling in general. With too many people the word modelling kind of provides them with a shiver up and down the spine when they hear the word, and I think the problem is that all of us deal with models, everyday, not just with our mathematical modelling, and that is, for example;the example I use with my students frequently is everybody has a model of how a car operates; my wife's is very plain. She models the automobile: the car operator by turning the key and toward the ignition. And it is very plain, every once in awhile she has to look at the gas gauge and she knows that it needs some gasoline and it goes on. The mechanic, despite some of our experiences, probably has a different idea of how the automobile works, and a little more in depth. I think the modelling that Les is talking about is critical and I think it is critical from one standpoint: it is not necessarily going to solve design problems. Good engineering is going to solve design problems. But modelling, I think, is going to provide a framework for understanding the system and for interpreting the data we get. It does not provide the answer. And modelling is not something that ends at some point, it is an overview process which is continually upgraded by new results. From the standpoint of the specific questions that we are talking about now, I think BOD is really providing a barrier for further progress in understanding what is happening with this RBC. The only thing unknown of that believe in Science are the conservation equations; it is what everything derives from that we do. You cannot do a material balance on BOD. So how the hell are you going to do any modelling using BOD then, if you cannot even do a material balance with it? You cannot get a diffusion coefficient with it I will promise you that. Or if you do I would love you to explain to me what it means. Then, finally, let me point out, the diffusion coefficient, if you measure it even with a specific compound, if you are lucky you are going to get plus or minus fifteen percent...if you are lucky. Those quantities in Perry's Handbook for chemical engineering, plus or minus fifteen percent is real good --- ----- that is with very defined systems. Still diffusion coefficient is important in this fundamental knowledge. When you get out on the field you are going to take that model and you are going to cut it down. You are going to go from the mechanic's model of the automobile to my wife's model of the automobile, but in true you are going to have some understanding, some fundamental understanding of what is happening in doing so. You are not going to be doing its modelling with exponential formula too where units are not even considered in order to design a system. Kinetic is always important. I bet you can count the number of reactions on your hand, the chemical reactions that we know of today, whose kinetics were determined at priory. Everything has got to be done empirically and it is done within a relatively small range so we can use relatively simple models. Make yourself something like what Dr. Grady was talking about which has three or four parameters to something that has maybe two, it could go up to a small range. Dr. Grady is looking for more comprehensive models. And that is admirable and it is necessary, but the lady that has a problem back here, she needs good engineering. That is as far as I see it, engineering is never going to be cut and dried. If it is I am going to get the hell out, and there is not going to be any fun then. Because of that is what these people are paid for, so I think there is a problem here thinking everything is black and white, and you know... That's about all I have to say.

MR. ATHAVALEY: Referring to which mechanisms actually controls there could be a lot of debate on that. One could study whether it is transfer from gas phase to liquid phase or the mechanism is in the interphase. If you open any kinetic book you will find there a chapter on retention time distribution, and all the constant data were determined on pilot scale or lab scale, are then taken into account when actual distribtuion of retention time is taken into consideration. I took a look on configuration of the plate or disc which are on display and the main problem there is distribution as referred to solid distribution or with wastewater retention depend on how this disc really performs, and for this reason there should be a combination of some integral relationship of kinetics or the practical application of kinetics are empirical, and combination of kinetics with fine distribution and the aid of particular species is important. I discuss that point the scale-up and design engineering with regards to kinetic reactor, with regards to fluidized bed reactor, are done on the basis of retention time distribution. The technique used for that is chromatography technique or photographic technique, so on so forth. I am not very clear at this stage how this could be applied to the disc reactor which is complicated and the reactors I have just mentioned. I am strictly talking from kinetics chemical engineering viewpoint.

DR. MUELLER: I would like to make a few comments on it myself. With respect to the model, I think academia for a long time has had a bad name with respect to model because the reactor you do not understand, and we tend to identify them with the utmost degree, that is where we make our money. We make our money from the research project. Why lick the problem? That is one aspect of the model, that is one aspect of the understanding the system. I personally think we have got to do that more and more. The more we learn on the academic money and the more the field will know what is going on with the system and they will ultimately design it better. That is one aspect. The second aspect is using your smart from academia and saying here is what we know of today of our modelling and here is what we all know about all the parameters we got to have in this thing. Now, with respect to glucose diffusion, I will be damned if I know how BOD diffuses, but I do use glucose because I have nothing better to go on. I think my kinetic parameters are incomplete and I got a lot of ignorance in the model, I am not sure how all these other kinetic coefficients readily do close, and I will be damned if they do not fit the model which uses glucose as a BOD diffusion parameter. Now, I get numbers for those kinetic coefficients. I cannot swear that those are the numbers of my need and I am not sure of the ultimate truth, but I can swear that those fit the models under different operating conditions, engineer to the other aspect, not the academic, not the research aspect of which I am a part. The engineering aspect, I want to evaluate that system this is the best I have. I have got a model, I have got imperfect information on it, hell, I am going to assume some numbers. Now, if I can fit that numbers for that set of data and I can get another set of data on the same waste on a different operating conditions, that is fit and I have another set of data on the same waste, with all those numbers I get on that data, the A, the K, the B, they will all be off, but I have got a fair degree of confidence that I can extrapolate this condition and that condition to that condition, fairly wide swing and still predict the point. Now, as an engineer with the best capability they have today, in my hands, from the academia and from anybody else, I can use that to extrapolate the other conditions: little question on how accurate it will be, but it is the best I got. Now, I think that is where we are today with the models. We can use it for design, we can bite the bullet, and bite somebody's coefficients and say if there is ever, we can apply this model for this case and for that case and so far we haven't said we can. There is a lot of room here. But we did it academically. The story today are selling equipment, Clow in the market and Autotrol in the market FMC in the market, and some others in the market. Independence, of how long it takes to solve his academic problem it is going to be a lot of work and a lot of money on his initial work that is going to get some solutions, just we have known today. So, it has got to be a marriage in it too. It has got to be a marriage of fighting upon accepting the best knowledge we have today. Based upon this information engineers can select the system on a market the best we have today. And we also have to say to ourselves that we do not know it all and we have to go out to get more data.

MR. OPATKEN: I would just like to bring up one point. We know the COD reduction at each stage. We know the DO level coming in, we know the DO level coming out. We know how much oxygen has been transferred. We also, with that same thing, have the mass transfer coefficient from it. I do not know where the problem is.

MR. HARRIS: I do not know anything about RBC's and so that it gives me advantage. I can ask a dumb question and perhaps get away with it. I have heard some comments and encouragements that RBC's will give a higher degree of performance than a trickling filter, so my question readily is why should that be since both units are essentially the fixed film reactor? I do not understand at all.

DR. BORCHARDT: I think I would go to the origin of that and ask that individual the question. I like to think of an RBC as a horizontal trickling filter myself.

MR. HARRIS: Rather than a fixed-film reactor.

DR. BORCHARDT: Sure, well that is what it is.

MR. HARRIS: Why should they in a sense be different?

DR. BORCHARDT: I do not believe they are different except that we are accustomed to recirculation in trickling filters and we are not in our RBC, that is the difference.

MR. HARRIS: The recirculation...

DR. BORCHARDT: The recirculation factor, yes.

MR. HARRIS: Is the one that makes any difference that is observed?

DR. BORCHARDT: That is the difference. We are not accustomed yet to recirculating. I think there are times when we should be in the RBC we are not, so far.

DR. CHOU: One recommendation as far as the research priority is concerned, we have talked about a problem of uniform methodology as far as measuring the organic component or components, and I do not know the percentage of nitrification required in this country. I think there is an important market and the ammonia analysis is uniform and there is easier thesis to work with so I recommend some research effort can be channeled into this particular area and that is my comment.

MR. DENNIS: In response to your question about the trickling filter I also do not know a lot about RBCs, but it is kind of appealing to think of an RBC as, or a trickling filter, as a secondary biological contactor and I think the reason why the performance in the RBC is better is because the reactor is longer on its plug flow system, whereas in the trickling filter you have about maybe six feet of filter stone.

DR. BORCHARDT: I want you to all realize that we are not up here because we know the answers to the question, it is we were selected because they thought we could listen politely to your questions, and tell you there are no answers.

DR. CHARACKLIS: I have a question too. I have been working with biological film system from a completely different standpoint, and at least from a different problem. Two of the problems we are concerned with:first, the problem is essentially energy lost, frictional resistance decreased, the film growth in the tube and secondly heat transfer resistance increased by the fact that there is a biolfilm inserted. Now, you know when we get into this problem you look into the literature for heat transfer coefficient and mass transfer coefficient for example, and you find that correlation in turbulence flow which is all you can use : correlations of these mass transfer and heat transfer coefficients always talks to friction factors in a flow system. Somewhat is easy, you have measurements to make too. My question is we use in one of our systems a rotating drum and grow a film on it, and the measurement of frictional resistance is simply applied a torque meter between the motor and the drum. It is a very nice indirect indicator of how much film is there but it is a direct indicator of the amount of frictional resistance at the surface, which should be closely correlated to mass transfer and in our case for example heat transfer. I wonder if anybody has ever tried keeping a torque or in fact whether the system was sensitive enough to detect these changes in torque on the shaft due to the biofilm on the surface of them.

MR. NICKLE: From municipal wastewater with four shafts and four stages, does it matter whether the shafts are loaded perpendicular to the shaft or actually parallel to the shaft from a hydraulic standpoint and process standpoint.

DR. BORCHARDT: Well, I think no. I loaded them both ways and actually the mixing within the stage is so complete, one time I measured this and I found that I got seven turnovers for each flow through, in other words, there was complete mix in each stage and it does not matter whether you put it in longitudinally or across the set of discs, if I understand your question correctly.

DR. MUELLER: I would like to discuss the dimple effect for the moment. From the actual, our initial workdown in that area with the Autotrol media, if they grow too thick a film, the densest part of the media and cut down active surface, and we can cut it down significantly with large film thickness. We do not cut it down with normal thicknesses of a sixteenth or so, you start the disc thickness of half an inch really hurts, you have go no surface area left for transfer of oxygen or substrate, but the principle effect of the function really of the thickness of your biofilm. The more thickness you get in the greater the surface area, but if you got a lot of biofilm thickness on it you just wipe it out like that. So there is got to be a joining of both effects, surface area by dimples made, and biofilm thickness. And that will optimize the system.

DR. BORCHARDT: Now, up to a certain point that film was completely porous on a microscopic basis, but it is that point that we do not know about that you are speaking of. Apparently, the anaerobic region on the dimple surface is wasted, and I think I am inclined to agree with you.

DR. BORCHARDT: Well, let us pass on to the next subject, because I feel that all of these things are so interrelated that it is very hard to separate them and we will just discuss this next subject and perhaps you will see what I am talking about. In any way, for Group III, we have selected a brief discussion here as RPM in Turbulence. You all recognize the fact that many of these have different surfaces, obviously turbulence is an important aspect of this whole thing. We are going to turn this over to Dr. Friedman from Syracuse University.

DR. FRIEDMAN: I can sum it up very quickly. I do not know, I do not think anybody else does either. But let me give you some ideas. All the models I heard today assumed the uniform liquid film across the surface of the disc and yet when I go into a laboratory or go out in the field to look, I notice that there is no such thing as a uniform depth of the liquid film over the surface at any point on the surface, or any two points. Now this has a lot of implication in terms of diffusion of oxygen from the gas phase into the liquid phase, or at least to the biomass. So that is one thing we need to improve in our modelling. Now, I attempted to do this a number of years ago when I went and talked to a resident hydraulic engineer, he was a darn good one; he scratched his head a little bit and went away and thought about it. He came back and said why don't you go upstairs and try chemical engineering. So I went upstairs and talked to the resident chemical engineer who was presumed to be good in hydraulic, and he sort of scratched his hands awhile, and said why don't you go and ask Dr. Lee. The person who produced a simple plate immersed half way it would not be too bad, but when we do not immerse it all the way or some other percentage we really have a tough situation. Now, I am saying from a modelling point of view to start with, we do not know how to describe the film thickness. Now, we got into another area yesterday, I heard was a kind of heat, how far apart should these discs be. Then the next question is what is the effect of the dimples on the disc, each manufacturer has a different set of dimples. At least that is what I call them. I am assured by one manufacturer that they have an absolutely beautiful computer designed that selected this particular configuration, and another one told me today we happend to stumble into it, but it is the best one. I am being facetious but I am really pleading ignorance. I do not know, and I do not know anyone else that really knows. Certainly this is going to be critical in terms of scale-up. And the answer is we really do not know, going back to playing games with industrial waste on how to scale the stuff. I think, and we have not tried it yet, nobody has seriously taken a series of three different sized scale-up models starting with two-foot, perhaps six-foot and then full scale and runs through a series of different parameters in terms of testing and looking at least obtaining some empirical factors that might be useful while we are waiting to develop our models to a level that we can really believe in them for predictive purposes, and not have to be waste speicifc and calibrate each model one on a particular waste. I would like to make one answer to Dr. Characklis' comment in terms of why BOD-5; because it is legally required, Bill, that is the only rational reason. With that, I would like to quit and open it up to the floor.

DR. MUELLER: I guess I do not agree with you.

DR. BORCHARDT: You will accept the first 120 microns.

DR. MUELLER: That has to be the case. It is the next four thousand microns that I am really concerned about. And under the anaerobic conditions. What you do, you shut off the whole channel for any kind of transfer. Once you clog it you shut it. Maximum 90% and probably half-way, you should have a good portion of your surface area, and we can account for that sort of thing, but the optimal design for all manufacturers is to keep it to a certain level so you optimize the area available for your dimples, the active dimple area if you can provide. That is not that big.

DR. FRIEDMAN: Jim, I am not arguing with you, but I would ask you, no I won't either, I won't embarrass you by asking which one has the best dimples. I do not think anybody can honestly answer that right now.

MR. MADDEN: Well, I would like to respond to the gentleman who asked the question concerning the energy usage and the biofilm thickness, and there has been a certain amount of work done by all the manufacturers in determining what the amount of energy required, what the torque of measurements would be with the increase in biofilm thickness, but there is the consideration that more study has to be done because the biofilm thicknesses change in character and if you get a long and stringy filamentous growth, you have one set of problems, and if you have a just a lesser stringy growth you have a different value of torque absorption, but none of the data that we have accumulated would be with the acceptable in what you are looking at because it is not as fine, it is not as tuned as you would like to see it. For example, we would do it with one meter measuring energy to the motor end and find out exactly how much current is used in that equipment. What you are probably looking at is something like the use of mechanical gauges and that like, and we did some of that work. It is not in depth to the extent I think you should look at, but we are looking at more of that, and I think other manufacturers are doing the same thing. Somebody else had a statement and I think it really belongs to the manufacturing pocket, concerning axial or longitudinal application. And speaking for myself, we would prefer longitudinal application as opposed to having it actually, and we do think you get better mixing in that way in spite of what you said.

DR. BORCHARDT: I will accept that.

DR. O'SHAUGHNESSY: Up until now I have heard all the things we do not know about RBCs and what is going on. If you look at the other things we did not talk about at this conference: activated slude. Over the past ten years, you realized that as this is the thing developed for building a lot of these plants and we are still merry go around, and they get messed up, but in the activated sludge system, you are able to change F/M ratio as long as you can to get enough oxygen in there. Now, we have heard models and we have heard diffusion, we have heard which dimple has the best effect and so forth. I have a simple question since we really do not know how to optimize anything from scale-up to whatever we want to do, and you want to optimize this design at when you get an oxygen deficient system situation in the first stage, where do you put variable speed motor or variable speed rotation in every unit so when we do screw up at least we will not be sued quite so bad. Any participant can respond to that.

MR. ANTONIE: We have looked at variable speed driver for a decade, and they are ferociously inefficient. We have earlier found that most mechanics are not good enough, and if you want to double your power consumption you can go into a variable speed drive. But it is very impractical. In regard to biofilm characteristics and energy consumption, that is whether it is a relatively filamentous or relatively thin layer, biofilm thickness control is a very important parameter, and our recent experience using air driven and air supplementing contactors, has shown a great deal of effect in keeping that biofilm as thin as practically possible and necessary, and you also want to make sure that you have got the full effectiveness dimple surface area.

MR. MADDEN: Regarding the variable speed drive feature. It may be worthwile for us to look at that from a viewpoint of study and something to consider in mathematical modelling, but one of the things we find that is a poor feature from an operational point is that to leave a trickling plant operator in the position of the doctor or the judge or the jury on varying the speed of a trickling plant, with two, four, five MGD behind him, gets down to be a very bad situation. Somebody decides that they want to run slow or faster and you can have a whole set of operational problems that we could probably get into in the next discussion area, but that is something we think that is a bad feature of going into variable speed drive or multi-variable speed drive within a waste water treatment system of RBC.

DR. BORCHARDT: Yes, that is going to be the next area, so any of you that are talking about operational problems please give the next speaker here a little chance, a little opportunity.

DR. HAUNG: Well, we have talked a lot about biofilm thickness. I think when you talk about biofilm thickness, refer to the thickness and say it is kind of misleading, for we are talking about how much of the active microorganisms which are involved in the anaerobic oxidation of the organic materials is in the wastewater, and all the time that people find no matter if the biofilm is thick or thin they all have the good BOD, COD or BOD reduction in the treated effluent. That is because no matter, you have the thin or thick one, even the thick one only the very top surface is useful. The rest of the deep inside are really junk because they are anaerobic and are partially effective in oxidation work they will do partial of the job and of which is not accomplished too much of the COD reduction ...think of these in terms of the oxygen transfer from the liquid or from the air phase to the sludge phase. Which one is more important in the external phase, or in the gaseous phase or the liquid phase? It is my personal opinion that the gaseous phase is much more efficient. I am not a chemical engineer or a physical chemist but in the laboratory when I use a YSI DO probe, first I have to calibrate the probe, I stick the probe into the water, and just shake it for a few seconds, then dissolved oxygen reach the saturation, so I personally believe the oxygen transfer from the gas into the liquid film is much faster than from the bulk solution into the stationary stagnant layer.

MR. JOOST: I do not get out too often, but I have got a couple of comments. As you know, activated sludge has been around a long time, lot of accounts have been made for mathematical modelling, still have alpha and beta factors, and all of a sudden now we want to pinpoint the RBC because this is going to be the panacea of all the problems. Well the RBC system happens to be just an ordinary biological process. The bottom lines is that the operator.. he is the guy that has got to produce the effluent, and no matter what we tell him by a mathematical model which is a steady state formula based on a given film thickness which really does not occur, they put on another alpha or beta factor. This is not going to help him operate the plant, that is one comment. The second comment was that dimple surfaces and air drive and supplemental air. I did not get a chance to speak this afternoon, we had a lively discussion. I would like to mention Jim Madden, we are the other company that tests bioshaft, that failed to get mentioned by the way. But anyway as far as the...way back when the Europeans developed, excuse me, commercialized the rotating disc unit they used a flat surface media; polystyrene foam. And when Autotrol and their development program came out with the configurated disc they you know, kind of haw-haw and laugh-laugh. But it is really not going to fly. But they do build a hell of a good pilot plant because we tested the pilot plant against the flat surfaces we had, sometimes the Autotrol pilot unit produced better than we did, other times the flat surface produced better. I guess it was about '71 or '72, and Ron you can correct me on this, when it first came...when you first came out with the manual about comparing the flat surfaces with the configurated discs, to say that the result of Pewaukee study demonstrated that the unit surface area was giving you the same performance as the flat surfaces, nothing was mentioned about RPM, and I think at that time you had maybe about a 1.9 RPM. The two-meter discs test at Pewaukee, you had about a 1.9 RPM and comparative to the flat disc surfaces, which had the same performance per unit area, but the flat surface was operating at 0.8 RPM, which, yeah, you can get the same performance but it took you a little more energy, a little more compact time to obtain the same performance. When you first came out with the air drive unit there were a lot of comments, in fact I even wrote a little dissertation on the physical disc design and I got a letter from Autotrol that is a no-no, that regard to the industries I had to back off on that, but that did come out with that patent, and in the patent the figures show that the air drive unit which can produce the supplemental air which can produce the same performance per unit area at a 0.8 RPM versus a mechanical drive at 1.6 RPM is just giving us the surfaces back which we robbed in the flat surface area. Now, flat surfaces can get the unit's performance at 0.8. When you got to the configurated surfaces you had to go to 0.6 to get that same process performance per unit area. The air drives, according to the patent now, you are telling me I can go back to the 0.8 RPM which is where we thought about it, and when someone mentioned about the dimpled discs and you know whether they fill up or not, I think Jack, when we were doing some studies at Ann Arbor, we both kind of agreed that the plane surfaces as probably the ultimate in surfaces.

DR. BORCHARDT: That is all we had.

MR. JOOST: You are right, that is all we had. And now that we can do the configurated surfaces whether it be channeled, dimpled or pimples, as I understand the latest one is the jelly roll, we still are trying to approach the same process performance per unit area as on the flat surfaces. Those are my two comments.

MR. GERHARD: Now we have had thanks to Bob Joost through the work you did at Ann Arbor and other places Jack, we have had I do not know if it is as much as or more than other manufacturers but we have certainly had one heck of a lot of experiences on variable speed driver and the effect on the standard, so-called standard domestic waste, and more so even on industrial waste which has some wild characteristics. We know and have the physical evidence to verify that the DO in the mixed liquor will change depending upon whether you increase or decrease the RPM or rotational speeds. The BOD removal or reduction process efficiency changes, the suspension of the solids coming in and/or generated within the reactor, changes with speed. The temperatures were found on the hot industrial wastes will be effected, the cooling effect of the disc is substantially changed by vibration when you have higher versus lower speed, the effect on large scale units as well. So from the standpoint of what was set up by Mr. O'Shaughnessy, yes, we have done this, I would expect other manufacturers to have done this, and I have seriously suggest to the basis protocol of this meeting this evening as to the EPA or other sponsored research that has to be one of the absolute musts. And then, Jack, the other thing that you have done at,or not you necessarily but the Ann Arbor facility has done, as you explained there, the NSF situation, National Sanitation Foundation or whatever its proper title is, can set up a program to compare various package plant manufacturers designs in so-called prototype size units to find out if number one, I presume they met their own claims and number two, how they met the identical criteria of other manufacturers say their stuffs. With this, would this now be a practical thing for the Army, if you will, for EPA who is by far the biggest funder of all the projects to the tune of 75 or more percent. It is perhaps some grants from the states or regions which will also be involved in that they stand to benefit by standardizing as much as they can between different dimples or pimples or whatever the case may be.

DR. BORCHARDT: Bob, the NSF tests against the criteria, the criteria is established by the industry and regulatory agencies and I suggest that you get the industry organized along with the various regulatory agencies to establish criteria performance and I am sure NSF will be very happy to test against that criteria.

MR. GERHARD: Well, I would like to ask Mr. Opatken also to express the opinion of EPA on the situation like that is there or was there or will there possibly be some considerations of that type of thing.

MR. OPATKEN: There is always consideration.

DR. BORCHARDT: We will have a general discussion after everything is over. Mr. Lagnese of Dunca, Lagnese and Associates in Pittsburgh is now going to discuss operating problems and many of the things that we have talked about the kind of plant operation. Once we have completed that and the discussion that goes with it, the floor will be open for general comments just as long as you want to stay here, so bear with us.

MR. LAGNESE: Before I mention about operations, sneak in here my own observations as we attempt to find the ways to fine tune our modelling and our predictions so that all our answers come out to be right. Is that I look back at the other process activated sludge and everything, I wonder if the RBC and a lot of the things today that were encouraged and innovated process does not. Maybe would benefit by the ten states standard mentality that I think all of us fought against, most of us I think of being the generation that wanted rationality and so on, but when you think back, I think of the activated sludge process has developed in the same genesis as this does, that they would have the same problems. But they had the benefits in a way as I look back now and I just sort of here scratching my head, is that some wise people just said there would be six hours aeration. We would put oxygen in and/or air in the terms of so many cubic feet per pound or per cubic feet of space and so on, and I guess what I am saying basically we got to where the activated sludge still has many of its problems but not problems I think. I think the things we are trying to do by just over-design and the RBC and to its credits, and I guess to the credits of the profession today and our very cost effective mentality that we are working under. We will fix up the design and says positively this is the thing that will work, and we are frustrated. I hear argument back and forth about things that I agree with but that is not going to do it. When we sit back I am sure the EPA would not like to, as they struggle to get everything down to the so there is no waste, but I think as we look back we have to keep that in mind. I think we have gotten to where we are with lots of things, clarification, activated sludge, whatever by over-design. I never see a plant, RBC or otherwise, that when it is half-loaded it works well. Things do not ever work well or they have problems when they get cost design, and that is my comment as an observation. Back to the operation here from an operating performance standpoint, there has been somewhat of an embezzlement attitude of the practicing profession to the RBC. To some the RBC represents a simple operating device that basically only requires the ability to actuate an electrical starter. There is no need to adjust, balance and/or orchestrate the many operating parameters and biological indices, as is required for the quote, other processes. I am sure that this must be the majority view of operating personnel manufacturers'representatives without benefit of a properly sized or an oversized RBC installation. On the other hand, there are many others who remain frustrated by the severe limitations to alter operating and process variables as a typical first generation RBC systems that I think we are talking about. In part, the frustration is due to the inability to rememdy an unforseen or unexpected problem of performance related to incorrect design decision or incorrect loading conditions. As justified as this is, I still suspect that a large part of the frustration as related to the limiting operating adjustments generally typical RBC systems, is that it does not permit that type of sort of technical superiority which we all see simply displayed which we smugly display in our dealings with again that other process. The RBC as typically made available does not provide the opportunity to exercise the true understanding of process fundamentals. There is no opportunity to have supplemental changing the FM or the CRT by altering the mixed liquor suspended solids or to changing the sludge return to control of solids flux problems, or to alter DO level to correct the sludge bulking problem, to increase sludge age to enhance nitrification, we go on and on. I am sure that an advocate of an RBC simplicity would argue that there is no need to speed up the operation of a biological system by the continuation of object because he has no bulking problem, no solids flux problem or no DO problem. Today again, the advocate of more operating control of RBC would counter quickly with examples to the contrary. It seems to me that the dividing line in the argument relates to the adequacy of the manufacturers' designs, the manufacturer, the design and the application. The past record is certainly not too commendable in this regard. The problems of under-design and facility malfunction are there in sufficient number to certainly provide a sound argument for the needs of more operating flexibility to offset an obviously limited design state-of-the-art for RBC. On the other hand maybe we are over a mountain and have learned our past mistakes and have now reached a point of better predictive design capability and can finally offer the RBC as a handoff operating process. In soliciting your views on this dilemma let me first provide my shopping list of possible operating controls which could be divided on RBC systems. Of course, one would be a course of variable speed rotating controls. Secondly the device to measure the rate of biomass on the RBC to both process controls and to protect structural stability, integrity. There is the possibility of step feed arrangement, and we maybe could fit into that recirculation by step feed if it is an easier design approach. Stage control which the problem of changing stages to fit a given condition and from my own personal experience have had to get down and remove my filaments by pulling out a piece of wood or a piece of baffle, and I suspect we could probably design something that could facilitate the change of staging by something simpler than what we have had to do. Dissolved oxygen and temperature monitoring, temperature in both directions which are concerns in industrial, sometimes being too high and nitrogen problem being low. Heat addition impossibilities, we have seen in once case an industrial application of the need to actually control temperature from getting too low. And, standby supplemental aeration options, I hesitate to mention that after being in the earlier session, but I think we would have to say that that is certainly an option possibly that could help control operations. I am sure that many of you could from personal experiences add other operating controls, would take away some of the ones I have mentioned, which might prove useful to a particular problem you have encountered. With that type of perspective, I invite your comments as to the direction which should be taken to the providing of operating control options in the design and manufacturing of RBC.

MR. VAUGHN: I have heard in the last two days a litter discussion concerning the micro-organisms which grow on the media, and I have gained the impression that it is quite important to observe the nature of the organisms which are growing, yet I find that most manufacturers are advocating the use of a quonset hut cover which at best seems to provide only access to the bearings on these units. I would like to suggest that research certainly should be done concerning the type of cover and the type of housing installations provided, especially when we look at scaling-up problems, because most of the research work is done on units which are more fully housed than what we find in the field, and certainly would invite some comment from the equipment manufacturers on that. Well, certainly on the effect that that cover has on the environment in which that unit is operating, termperature of winter and summer, and from an operator's standpoint if it is important for the operator, which I assume it should be, since it is important for the researcher, to observe the film that is occurring on these units. Then I ask the question how can you do it with the present housing? How do you expect that operator to even be able to see what is happening on his RBC when his access to the unit is only at the bearing ends, and this to me for an operator control would appear to be an important feature.

MR. HYNEK: I will answer that question directly. I will start it off from a practical standpoint which has been realized by Region Seven of the EPA where where a lot of the RBCs started off with buildings, at a point in time when covers were not available. And at Region Seven you have the State of Iowa for example, cold temperature in winter time, and the operators they have to grease the bearings, they like the idea of being able to walk around between the shafts and so forth. While Region Seven no longer advocates buildings, high humidity, corrosion, coupled with OSHA requirements, lighting requirements, heating requirements and the bottom line was heating. One plant spent a thousand dollars a month to keep that plant at the near normal temperature. With respect to the cover in terms of access for an operator, let me go back to what I feel are the basic tools an operator has to know if his plant is operating correctly; eye, ears, nose and smell, he can look inside the door and that biology on the first face of the reactor is a reflection of biology that is occurring in the reactor not exactly the same by any means; there is a different hydraulic profile, different hydraulic exposure, but it does reflect changes just as well as what is going on inside. So, as eyes can tell day to day changes, and this has been common experience. His ears, if he does not hear any noise, he knows something has stopped because the drive motor on the mechanical system is there. On the air drive system there is a kissing sound from the water moving with the bubbles and so forth. He knows, just like that, he knows there is something different if he is there day by day. So this is the basic virtue of the RBC process, from an operator's standpoint. His own senses can tell him very quickly if something is wrong. With respect to the atmosphere within the covers, there has been detailed technical data developed early on, is there enough oxygen diffusion in and out. There has never been a proven case of a lack of oxygen. The closest in a quantitative term is possibly five percent differential, and I challenge anyone in the audience to do it any better than it has been done. For building however, we have had documented cases where the oxygen content in the building because they hook the light in the building with the fan. When this light switch went off I think, and the operator went home, the fan went off. Ten percent of oxygen in the building; you couldn't light a cigaret, you could not maintain normal feelings, that you were in a good oxygen environment. Performance however, not measurably different. I hope I have answered some of your questions. And you can take samples from the radius of the media, look at it under a microscope, and see the microflora and monitor that on a daily basis, if you really want to get into the fine art of seeing the subtle difference.

DR. FRIEDMAN: Could I ask a question. That is fine, Bob, all these signs of what we do, but the question really is suppose you find something wrong then what do you do?

MR. HYNEK: It is a good question, and it has been a common question. Well, if there is something wrong in the sense that he senses that there is something wrong, we have found that many basic unit processes either up front or behind have been a factor. We have several manufacturers present that I think we can liken to pioneers because we are early in this game, and so you try to answer these questions. We have found from extensive field evaluations of all equipment, full scale, that there are problems up front with respect to solids removal. There are problems with solids removal after in the secondary clarifier because the operator does not want to pump properly, he does not have really simple automatic equipment to put her on a timer basis. The digesters cannot get the anaerobic digesters up to speed: they have waited for extended months of time, they might have local problems. One recent investigation ocean outfall, when the tide comes in it backs and floats the head works, and they have to elevate the waste water to make sure it will go through the plant, so there really are a lot of serious problems outside this new black box, and I think we are getting hit from all corners, from consulting people who are sensitive to their customer, from our sales representatives who are sensitive to the consultant, and the customer, they tend to panic. I think they have got to slow down and cool down and realize that you can expect this panacea concept which perhaps the early-on RBC created, and well we are now dealing with biological phenomena, bug is bug is bug. It can do so much, but mechanically these other systems can impact on the biology, and they have to recognize that.

DR. SMITH: The Army has been recently mandated to cut back their enery consumption by twenty percent by 1985 or so, and in lieu of the fact that there does exist a common problem inherent to RBCs cold weather winter operations and due to the fact that we are trying to push solar heating, solar collection, etc., I was just wondering if there has been any activity in trying concretive efforts to make the covers actual solar collectors. I am not saying make them clear because you would have algae growth of course, but maybe some sort of mechanism. Painting them black would be simple except during the summer they may overheat. I was just wondering whether that would be an obvious and simple way to solve the energy problem, feeding, and also cold weather problems. Somebody have an answer to that?

DR. FRIEDMAN: I think the main reason for utilization of RBCs in the country is the energy utilization and the reduction of energy consumption. The EPA was wise enough to fund the energy plant in Maine which incorporates some of the panels over the discs, the first totally energy free plant in the United States, using anaerobic digested gas to heat the building and deriving its energy source from the solar panels. At this particular time we have discussed many things with regard to RBCs in operational problems but I think the topic of discussion of why RBCs are being used are basically its simplicity of oepration, its guarantee of a flexibility should be incorporated in the system, and the capability to remove BOD at an extremely low horsepower. We are discussing a lot of problems but I guess maybe one of the topics of discussion should be optimization of total energy uses in biological waste treatment, and I think that the statement that was made regarding to power, consideration of the Wilton,Maine Plant potential future studies with regard to energy utilization using RBCs should be considered.

DR. YU: I have a question to be directed to the manufacturer. I do not know whether the design from the different manufacturers have a standardization in terms of all of their equipments or not. Based on the past experience, some of the big companies due to business reasons, want to shut down certain divisions. So when you shut down the divisions field service is no longer available to what is exactly the duration of a biological disc, so in case you need it and it happened to be that division of the company that was shut down, where do you get it? And, you know, this has more interest than those kinetic constants. When you need it, K does not mean anything.

DR. BORCHARDT: That sounded like a statement and not a question.

DR. YU: Well, I say the question is directed to the manufacturers, whether they have a standardization or not; when you get it from the A Company, if the A Company shuts down, you get it from the B, and you do not expect property owners have to take care of their own problems.

MR. ANTONIE: With the standard RBC components we originally designed and were marketing in '72 and since, have sort of become the standard at least from the standpoint of shaft length and tank and so forth. And I think you would find that any plant that has been built since 1972, you will be able to find two or three RBCs that could fit that structure.

DR. BORCHARDT: That is, if you go out of business, as I understand his question.

MR. ANTONIE: Incidentally, people that have gone out of the wastewater treatment business did not do so because of competition, they did so because there is no money being spent.

MR. HARRIS: I asked the question earlier which was not answered, so let me rephrase it. Is there a fundamental operating advantage of the RBC over the trickling filter?

DR. BORCHARDT: Well, we heard your question the first time, and there were a couple of attempts. We have not guaranteed an answer to every question, you understand that.

MR. HARRIS: Yes, I understand. If the answer is no, then I would point out that there was in the past couple of days one paper at least given comparing the energy requirements between trickling filters, RBC and activated sludge. The RBC as I recall was somewhere between the trickling filter and the activated sludge. Now, if there is no particular cost advantage for the RBC, and there is a definite energy advantage for the trickling filter, why don't we have a first national conference on trickling filters?

DR. FRIEDMAN: I think your reasonings for standard rock trickling filters are no longer in vogue; it is terribly expensive to handling rock and be sure it does not crush. There is also a difference in space requirements for the two; precisely they are certainly space limited. The RBC installation would be smaller.

MR. HARRIS: Yes, the standard filter is on the rock but there is another type that does not use rock.

DR. FRIEDMAN: That is a packed bed filter. You did not ask that. That is much more expensive. Do you agree with that?

MR. HARRIS: I did not have the basis for agreeing or disagreeing.

DR. BORCHARDT: Yes, I would have to say that if you use plastic media, the trickling filter would be much more expensive.

DR. WU: There are two papers that will be presented at tomorrow's session to talk about economic evaluation of all kinds of systems--activated sludge, trickling filter and RBC. I think the authors may answer your question better from their paper presentations.

DR. CHESNER: I want to get back into the operation of controls a little bit. We had some discussions this morning on supplemental air and rotational speed, and I am not quite certain what the exact effect of each one is, or what each one does individually. I happen to think they both probably do the same thing if you add air to increase rotation of speed, increase the turbulence, increasing the net transfer. In terms of any operational control in the most efficient manner, and I am trying to question out the manufacturers at this stage, it would seem that having supplemental air standby supply and mechanical drive, is probably the most efficient way to go about operating the system because you do not have to run that air to drive the system continuously and you can only put it on and put it on when you feel it is absolutely necessary from a profit control standpoint.

DR. BORCHARDT: Well now that I think is a statement that you want a yes or no to.

DR. CHESNER: What do you feel about that?

DR. BORCHARDT: Well then what would you say, yes or no?

DR. CHESNER: There is a difference of opinion in terms of whether we should increase or whether the operational control should be an increase in rotational speed or should we have a natural system driving this biodisc by itself and supplying it simultaneously?

DR. BORCHARDT: I have a feeling that up to this time the air drive system has been used because it was thought that the air drive system has advantages which would recommend it. I think there is some areas which the positive mechanical drive might suit the situation better and I think this is an engineering question which has got to be answered as the problem comes up. I am not sure that answers your question either, but I do feel that is an engineering decision. This is not always so, one way or another, you see, so yes or no is not the answer.

DR. CHESNER: I think the real issue here is energy though.

DR. BORCHARDT: Well you can make energy one issue but you have to have the thing run. It does not run when you are saving energy. That is not the answer you said.

DR. CHESNER: The answers with the mechanical drive, you are probably going to keep that system running. The decision is to add an operational control but you want the operational control, whether that would supplement the air. That seems to be the only control unit that you have. Supplement air in increasing rotational speed and it is not an intention that operating the system with mechanical drive and having supplemental air drive as stand-by is the most energy efficient way of operating the system, energy efficient way having some control over the system.

MR. LAGNESE: Jack, I do not disagree with that. I think in the area we are talking about, of what should we do with operational control, I think it is a good question and I, the way I look at it is that we are going to continue to try to pay for performance that we are not going to give more than what we need. I think we have got to talk about operating control like that. But I think the alternative to make it bigger and you do not need stand-by supplemental air or fast procedures, but I think for some reason we have gone into this thing that we are going to try to come as close to it as we can. From what I have seen we had better then put in operating control and I think I would buy your logic that we are talking about one of the other and I think I agree with you that maybe the supplemental air is the easy way to go.

DR. BORCHARDT: You mean you want to add both the mechanical drive and the supplemental air in case you need it?

MR. LAGNESE: That is right. This is an operational stand-by for the possible that we cannot pay for design or waste conditions to the exact conditions. I think the alternative is to make the system bigger. But that is not the way it is going.

DR. MOLOF: I would like to talk about this idea of the dimples and all that stuff. But relate that back to the manufacturers' getting together to create a standard. And in the drinking water field the manufacturers of home water treatment units formed an organization with a name called "Wackysack". I hope you guys did better than that. And what they did was they did not let EPA run it. A fellow from EPA was on the board, they wrote to the National Sanitation Foundation and they remained a private industrial group setting their own standards for quality. And I think this might work very good with the group here. The fellow in EPA was Frank Bell, the National Science Foundation with Dr. Nina McCollum. You can find out and get together and do this thing but you have a model that has been set up already, and they really set tough standards. I would like to talk about the dimple effect on area because I have had the haunting nightmare that we are playing a numbers game with area. And I think one of our problems is that engineer is going to get involved with is relating pounds of BOD and I know you do not like that word, some people do not like it, but what is the pounds of BOD per thousand square feet per day? What is the square feet on Autotrol, on anyone's unit when there is a half inch of biological growth on it. How do you explain it? This is the problem that the manufacturer is going to have to answer because I do not think any of them can define what the effective area under actual operation is. I do not expect an answer but just accept it as a challenge for me to them, if they cannot define the actual operating area. Now maybe that is why our design engineers are in trouble. They are told of an effective area and they design it and they overload it and they need supplemental air. I would like to bring you up the fact that speaking on operation, really gave a very good viewpoint to the fact that the operator has very little control which means the design engineer has almost full control. You poured it out correctly in an activated sludge plant but they have enough valves that turn and buttons to push. The design engineer can just give them to you and you can operate the plant. You are absolutely right that this plant you lose that control and I always search for ways that people can adjust to conditions that were unforseen. I think that is a very good point and you ought to follow it up. But that does not make that one comment to help Mr. Friedman. About 1975 I put in a proposal with somebody to study what they think between 3-foot and 12-foot and they thought I was nuts.

MR. MADDEN: I would like to comment on some of the things that Bob Hynek pointed out concerning the operations of difficulty. We have found in the majority of our operating plants on line and discussing the various operatives that many of the problems in the RBC system do not originate in the RBC system. I would say upstream of the RBC in terms of pre-treatment or advanced treatment in terms of clarification, infiltration or whatever we consider as the system. And we have found that upon looking at the problem that it is not something where the design engineer took it to account perhaps of the addition of the supernatant coming out of a centrifuge or something of that nature, or a mis-design in terms of dumping this centrifuge supernatant directly into an RBC system without passing through the primary clarification. But when we looked at the problem and we found operating facilities and why they have not been many, and they have not been terribly complicated, we can in almost every case trace them to problems such as these older problems and the like, all have their bases in either upstream or downstream treatment and really not in the RBC system. Concerning trickling filter, it has been our experience, we have one of our people operate a trickling filter and RBC plant, and the general comment has been that the trickling filter is the rock filter that we have experienced has not been able to achieve the degrees of treatment that contemplates with the RBC biological contactor. And while there are others who may differ with this viewpoint, the fact remains without high rate trickling filter or high rate recycles, you do not get the degrees of treatment that you do with the continued ones through RBC systems. I think that might help out the gentleman from Shell Oil regarding his question about RBC and trickling filters.

DR. BUNCH: I would like to make a plea for all the little villages and hamlets of 10,000 or 15,000 people who have a $50,000 per year budget to run the fire department, the ambulance, the water department, the sewage department. We stand here tonight and talk about your operator's experience, and say we need more operators. I would hope that we can get the RBC plants to the point that you do not have a round-the-clock operators.

Any time you have to have a round-the-clock operators, you are going to triple your cost and labor, and labor is going to be one of the big items in the future. And I would hope to think that the RBC system could cut down on the number of operators. And the second thing as far as energy being all that important, so important that the local scale, because on a national scale the amount of electricity required or the energy required to operate all the sewage plants in the United States, there is a very few percentage of that total energy budget. In fact, it is about...to operate a sewage plant would equalize leaving a 60-watt bulb on 24 hours a day per capita.

MR. DIAPER: I have not got that much experience on RBC but I do have experience at another location with a device employed for the other purpose in wastewater treatment which we are pioneer, for the microstrainer. And even through similar problems that you are going through with now with the RBC's, and I agree with Joe about the need for conservation in design and you have to have a safety factor because you cannot cope with these unprecedented contingencies that occur in the sewage treatment plant upstream. But these as with the microstrainers are fitting into the variable speed drive. And first of all we had it manually operated and then we alternated it and we controlled it by the head differential across the strainer. Obviously you are not going to get head differential across the RBC and maybe you will, but surely there is some parameter on the RBC that you could use such as film techniques or load or viscosity to control the speed of the motor automatically. I certainly think you should give the operator that opportunity to control the speed manually even if you can do it automatically. And the second thing also arising from my experience with the microstrainer is after being pioneering for several years, the standards will be developed for design and they will be developed from a simple experience based on ten states standards. And I wonder no one discusses it as this discussion has pointed out that there would be a better control method, as it has been pointed out, it would be a way controlling the surface area for giving flow if the 10 state standards would be included in the design section of the RBC.

DR. BORCHARDT: The British have a large DO electrode that they use to control submergence of surface aerators. Such a DO electrode might in the first stage of the RBC be used, to be used to control speed which would possibly help us with some of our low DO situations. But as you heard from Autotrol regarding the use of variable speed drive the primary results are extremely expensive and so that is one of our problems.

MR. DIAPER: Well I think you are going to get towards the condition that Joe is pointing out that you are going to have to consider the safety factors in the design to get away from these operating problems.

DR. GRADY: I want to make one very simple comment. Of course some of you may agree that I have made a lot of simple comments, but anyway, you now talk about factors of safety. One way to handle uncertainty is what is called a "minimax" design and that is you take the worse situation which represents maximum size installation but then you optimize the cost of that system in order to minimize the design for that particular circumstance. And I think right now given the uncertainties that we face on all of that parameter value, but this is really the approach that we should all be talking throughly about these systems; however, no one knows how to put in good operational controls. We do not know what is going to happen from now on.

MR. GERHART: In an example of an installation at a petrochemical complex that faces that given problem of control variability and flexibility on a plant that had not been built yet, you could not even have a pilot plant study, let alone model or anything else, and they wanted a process guarantee which, of course, it is absolutely not but what they said you could offer us in the way of flexibility because we will not spend more money than we have to. And the net result was that two-thirds of flexible system, one involving in an equipment with six shafts in a single flow stream. Now they fit a parallel channel so that they could bypass any third or step feed as they desired. This is a very simple method of control, not a very costly method of control. Second thing they did was to fit in the outflow in stages one and two. The DO flowed in stages one and two separately would indicate to the operator that if it dropped below, and I do not know where they set it, 1-1/2 two parts of water, but certainly no less than one, that if he dropped below that level in any of the two stages, then he should increase his RPM. But those things as we know from the personal observation of the pilot large scale systems. But the big thing was that it could load cells under the three end bearings on the first floor shaft in the train. But if the weight of the total operating shaft exceeded some value and I do not know what value they set. If we set five BOD average with 500 or 1,000 parts per million we thought at least the work done by Belgium's study. Any way they put the load cells into the free end bank of the first four shafts. If any one of them exceeded a certain weight, it would then be indicative of a thick biomass film if you will, that they should then speed up or change up the RPM, or one way or the other I suppose. You know that the biofilm can be thickened down by increasing RPM or shearing forces.

DR. FIREDMAN: Do you have estimated the additional cost of energy to the overall facility?

MR. GERHART: The price that I quoted was about ten percent. That is what they told me. The engineer who designed the systems provided this information but it is not definite.

DR. FRIEDMAN: That seems like a very marginal price to pay for that. But I would like to point out we are really not talking about factors of safety we are talking about what a Professor of mine used to call factors of ignorance, and that is what we are really addressing.

MR. GERHART: Well when you talk about a factor of plus or minus fifteen percent, man that is the difference between a successfully operating plant and a lousy plant. And in terms of the operator that discussed his problem a little bit ago, if he has got the local enforcement agency on his back because he is fifteen percent over his permit, he does not give a stick or damn about the K-factor or anything else. He says what can I do about it, and one little fellow in Iowa, who did not have our installation, he got so damn desperate he went out and got some garden hose I understand, poked some holes in the damn thing and tied up to the various flow systems. But by gosh he got his thing up to make a difference.

MR. HYNEK: I concur with Bob Gerhart's remarks regarding the low cells. We have done a lot of work with it. The technology is all here, it is out there you can buy it. It can be done with simply a mechanical hydraulic jack operation, or you can put it on the electronic system and read it off the panel. So that question has already been addressed and taken care of. It is a little tough to sell it, because they do not realize the importance of it. One of the other things I would like to mention that has not come up here, is that we have had experience, I will not call it bitter, but recognize that the hydraulic distribution, when you have a central distribution point and you have several tanks through which the wastewater has to flow; there has been evidence of poor engineering in terms of distribution of the flow. And you will overload one bank of shaft and unload another, and if particularly you are into nitrogen and you have a poor operating plant, potentially. The other thing is, we have maintained the program with staying in contact with the operators, and we are hearing all stories from the standpoint that the little localities will not pay the buck. So the guy that will run the plant correctly, puts some dedication into; time after time we have cases where we, in fact there is one on the record where the State Department of Environment Control actually took the man to court for falsifying the records. He was not qualified to run an anaerobic digester and he was fudging all the data to make the plant look good. Another case, one of our plants ran four months with no attention: as testimony, there was never any snow tracks or shovel marks through the shafts for four months, no lubrication. He was fired. So we have had problems at all levels, it is not just the RBC.

DR. BORCHARDT: The last subject, Mr. Opatken is going to talk about research support funds. I know you are all dying to hear it.

MR. OPATKEN & MR. BASILICO: We started out with a sub-chairman section and a couple of committee members and we met yesterday last evening. And the first item on the agenda was my committee members; how I wanted them to get to the committee, that suggests me immediately changed the subject. So I am left here by myself. When etc., etc....we regret to inform you that...this is not standard. It is not a form letter but it does occur. But there is a reason why it occurs. People submit a good research proposal, and the proposal will be on a subject which may be reaching the top of a program plateau. It is an on-going program; a program that has been developed, objectives have been defined and there is movement towards reaching those objectives. You come in with a proposal. It just has a very poor chance at that particular time of being funded. You have filed at a time with high visibility. But what you have to begin with is the fact that what you want is a program that is on the rise, a program that will become visible, then your chances are much better. I offer that just as advice because that is about the way the game is played. The pie is cut up and it is cut up on the basis of priorities. I will say that if you have a project that you feel would offer a significant improvement, it certainly would again be considered, but I will not define significant. Now what if you do have ideas set forth. The best method is to contact the Wastewater Research Division. Get in touch with the person who is involved with the particular area that you are interested in. You make that contact. It is nice then to discuss what you have to offer, the data that you have and we like to then set up a presentation by you to a group of peers for evaluation at that particular time. Then a project is then evaluated in an informal manner and it is then encouraged or discouraged without going through the formality of a full-plan proposal. That is one of the fine methods. Another method is to submit your pre-proposal. But again I say, you know what you are working on is the program priority level. Well, the Office of Research Development in EPA has recently come out with an exploratory research program for university professors who are interested in research. Basically that kind of work is going to be the peer review of proposals, and the reviews would be based on certain criteria. And the Ad Hoc reviewers will be selected from a group of people that have been nominated. Some have been self-nominated, some are still being nominated. They are the experts in control technology, industrial, municipal areas, and so forth. The mechanism for funding has been organized and the information to be announced by some EPA research pamphlets and flyers. I have got a draft right now it lets you know how the system works and the type of reviews and the type of various applications that we process and so forth. The funding actually have taken money out of one pocket and put it in the other to pay for this new exploratory research program. The thing that got this research program initiated is that during past four or five years our program has been criticized for lack of exploratory type of research. Somebody said the EPA was only interested in applied research so that we really did not apply our program to funding to University type research. The EPA administrative officers now have decided to take the bull by the horns and apply the money from research project to set up a research grant program that does give the universities a chance to come in with some wider range type of proposals that are related to various control technology we are interested in today, and I think that gets us to the point that you may ask me how these priorities get set and so forth. I do not know where I think what industry does, you do, or you try, and as I would like to say to you I would like to get as a program manager for research. I am worried about the fiscal eight-two and so forth.

since we have to run the program that addresses the immediate priorities that were set two years ago and I guess what I would like to see is some kind of perceptible proposal relation to research needs at least come out of this symposium. Maybe somehow the workshop chairman can summarize the results of discussion in terms of research needs, to get some good research proposals from this group and eventually as a result of panel review and so forth. We have to make a decision for what subject should be funded; fund a project that is going to emphasize a bigger or better model. My impression is, we have got it from the industries and from the university, and so forth. From the proposals we should know what are the real needs, what are the priority needs and they could help us in the formulation of a very useful program for wastewater treatment. In fact, we need some more research and as I say it is supposed to be made on RBCs. However, we first have to define the objective: why do you need to do that. It takes all kinds of stuff to set up a program just because of the kinds of procedures required. I think it has got to come from outside and needs your input. That is the crisis of today. If you have questions, I will be glad to answer.

DR. BORCHARDT: I guess we are all from without, is that what you are saying?

MR. OPATKEN: Well, I have to say so I guess. Our program direction now aims at the operating programs, the construction grant programs, the regional offices guaranteeing what the states, what the consulting engineers, the priority program nails down the toxic pollutants control technology and our research program is geared toward the solution for toxis waste disposal. In fact, our construction program now has two or three hundred RBC plants and you know they are having problems and maybe they ought to look into you know, better control parameters and criteria for design. So our research program is more geared to obtain input from outside. Dr. Borchardt suggests to institute a research committee. For instance the research committee on wastewater treatment research and it has been a problem in our program from the regional people and hopefully we plan to have a completed strategy from this committee. And we do plan to send the proposal out for peer review. As we would like to send it to ASCE RBC Committee first because the Committee has already taken the position to study the RBC research needs. I think that could go a long way in helping the guy in the EPA research center.

DR. BORCHARDT: Well, you chaps all realize then that that suggestion you made for research is now going into the computer list as a priority project and we know the EPA will fund grants in the future. The group did not realize they were involved in that sort of thing tonight or a lot more of them would have gotten up and had a say what is happening.

MR. BASILLICO: No, I think the nature of this conference really has brought up a lot of good discussion on research needs in these areas, and most conferences do not end up like this I guess.

DR. BORCHARDT: Well, are there any comments now on this idea of research? Anyone else have research funds that need to be spent?

MR. VESSIO: As far as my involvement with the biological contactor is concerned, most of my experience and expertise has been with the application of mechanical surface aerators in municipal industrial processes and sitting in the back of the room I was listening to the various discussions tonight and listening to the other people who were conducting this particular meeting of today. I see a very close relationship between the development of the RBC and the history of the mechanical surface aerator device in this country. As a matter of fact the RBCs are at that position that the mechanical surface aerators were marketed in the early 1950's. Yoomans Brothers which is now part of Clow Corporation were the first company to market on a new scale mechanical surface aerators in this country and did a very admirable job with it. Mechanical surface aeration device just like the RBC's in Europe and in fact they utilized in 1929, but were never really introduced into this country until the early 1950's. And the first aerators in this country were actually marketed on a license from a person in England. And the expertise was infant just as it is here with the RBC. And it was not until 1965 and by this time there were two or three other manufacturers of surface aerators who got into the scene just like you have here with the RBCs; you have some initial works done by Allis Chalmers, a lot of works done by Autotrol; there were companies like in Envirodisc that came on the scene, TAIT bioshaft, Walker process and so forth, each one making different claims, or claiming some feature of superiority of their particular product. In 1965 at the TAPII Convention, Yeomans Brothers developed a standard for testing mechanical surface aerators and many equipment manufacturers utilized that method. That particular procedure is now a part of standard method. It is also written up in the manual published by the water pollution control federation. The whole point for what I am saying here is that why can't there be a standard for determining efficiency for the various RBC designs extensively used today? An interesting paper today gave the right terminology which reference was made to closed pack media versus open pack media, which caused some lively discussion and I think that this is extremely important. Because the little bit of exposure that I have had with RBCs indicates to me that it is going to be as it is right now I think, or it is certainly training to be, the equipment of choice because it has several advantages of operational and power-wise and so forth, from any of the processes that have been used in the past and are even being tested today. We have found some of the wastewater treatment processes which are not really economical. I would suggest to the gentleman from the EPA that the true consideration be given to testing for efficacy the various designs extent and being offered today by all the manufacturers, whether dimple flat sheets, jelly rolls, whatever. That these units be tested side by side on similar type of waste and conduct an independent determination as to what designs seem to have the optimal efficiency.

DR. BORCHARDT: Very fine. I gather from what he said that there is hope for the future, that there are problems to be solved, and hopefully there will be resources to solve the problems. Mr. Opatken says yes, there will be. Well, we are now a half hour overtime followers and unless there is some violent objections I am going to call the meeting to an end.

APPENDIX A

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APPENDIX B:

FLOOR DISCUSSIONS AFTER EACH SESSION

Session 1. GENERAL DISCUSSION
Presiding: J. I. Abrams
Department of Civil Engineering
University of Pittsburgh

MR. COSTELLO: The question was did we inspect the preconstruction sampling survey for the fourteen facilities that indicated variations in wastewater loading.

DR. CHESNER: Answer is we did not.

MR. DABROWSKI: The question was, was there existing on paper the location of each of the individual facilities.

DR. CHESNER: I did not but if you request to me personally I will let you know.

MR. CONNER: It is very dangerous to take energy data especially in terms of energy per unit flow and just use it with blank checks. The reason is you have to consider the efficiency and the performance of that individual facility with the energy utilization. Some of the facilities we deal with energy survey that there was such little data available concerning energy requirements especially considering the importance of energy these days.

DR. CHESNER: I would say that the large variations between RBC and trickling filter even though we did not have a lot of data, tends to indicate that these numbers were pretty accurate in terms of differences in energy that RBC were less energy efficient than trickling filter but certainly more energy efficient than activated sludge units. In our field testing we will be getting more information on RBC units in terms to establish relationships between flow and energy utilization and between efficiency and energy utilization. However, we still have that gap in terms of trickling filter and conventional activated sludge data.

MR. DABROWSKI: Your company has been asked by EPA to evaluate RBC wastewater treatment process. What is your idea to proceed on this important assignment?

DR. CHESNER: I would say that depending on existing applications and sort of, get a try to score that question. We would certainly listen to all manufacturers and right now in the process of designing an RBC facility to treat seepage which might be the first facility I think in the country that is eventually doing that. We would certainly entertain each manufacturer and listen to what they had to say and them come up with the best judgment. I do not know if I answered your question satisfactorily.

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MR. HERKER: The question was where design curve indicated a percent removal of total BOD and what fraction of that total BOD was soluble and what fraction was insoluble.

DR. CHESNER: There are different design methods; some utilize total BOD and some utilize soluble BOD. That top method just utilizes total BOD. Unless you can predict the soluble BOD and your system it really comes out...There is not a great significance in using a total BOD or soluble BOD. You have to be able to predict the soluble BOD and if you can you have probably been dealing in a more accurate method.

DR. YU: From the data you have presented it is quite difficult to generalize the plant performance as a function of the process operating conditions. Can you comment on that?

DR. BANERJI: I really cannot from the data we have. I cannot draw a conclusion but the data...we are not addressing to that question right now but maybe eventually it will come that...But that does not give you a clear picture because that is not the only consideration. But this is just a sample of data we have but not more data available and you have got to look at individual cases now, than make a general statement. But on this BOD you cannot use it, so I do not think you can generalize it depending on the waste you have got.

DR. YU: I have some difficulties in understanding of your data presentations in one of the table showing the plant performance. Can you again discuss that?

DR. CHESNER: There are a number of columns on that table; one that I did not get into was effluent criteria problem and those were problems that the plant, at least 50% of the time is not meeting its effluent criteria. The other problems or some of the reported difficulties were associated with instability in the system.

Session 2. PROCESS VARIABLES AND BIOFILM PROPERTIES
Presiding: R. Neufeld
Department of Civil Engineering
University of Pittsburgh

MR. ATHAVALEY: I am asking the question from the viewpoint of chemical kinetics as opposed to mass transfer you have presented here. As we all know the depletion of initial biological oxidation demand or chemical oxygen demand is the final goal that quality should meet. Now considering the equations that you have here, are there any means to correlate this oxygen transfer to percentage reduction, say for instance BOD.

MR. KIM: No. This is solely physical factors in clean water test. There must be some limits of this system which it can transfer oxygen into the liquid film in the bio-reactor. So this does not include in our study.

DR. FRIEDMAN: This disc is heavily loaded, there indeed is going to be some sort of a correlation between the factors that he has presented. For actual RBC treatment system I do not anticipate that you would see any difference.

DR. MOLOF: In an activated sludge system everybody sells an aerator based on pounds per horsepower hour, as well as detention time, F/M, etc... In the disc field all we ask is a square foot area. We do not ask the question what is its aeration capacity. We are trying to introduce to this meeting the concept of saying, not only how much square feet area you have, what is the aeration capacity of your units on certain pounds per horsepower hour. I believe we need that incorporated into understanding these systems a little better.

MR. MOORE: In summary do I understand your relationship to say this, that the oxygen transfer capabilities are reduced as the space between the discs reduces, that is the smaller the space between disc media the less efficient the area's capability. In addition, what you are saying on heavily loaded systems, the thicker the film gets obviously the less efficient in terms of later on in the aeration capability. You get the distance between disc surfaces decreasing.

MR. KIM: The K_{La} is per unit volume and it is related to the volume of the reactor. This is very simple system in which you have liquid and air, and you pump water up in the air and bring it back. Again, this is a very simple system which you pump up water in the air and saturate it to some degree and bring it back. So the "F" is a function of reactor volume.

MR. MOORE: You are saying that it does not relate the distance between the discs but it is a function of the reactor volume.

MR. KIM: It is related to the distance of the disc. When I derived this equation, the only factor left in this equation is distance between the discs. But the above mentioned factor is somewhat related to the volume of the reactor.

MR. BACHTEAL: Would you keep the mass flowing rates on COD or BOD. I missed it when you went through it.

DR. ORWIN: In terms of BOD we are talking, three point six pounds per day per thousand square feet and seven pounds per day per thousand square feet. We tried to bracket some of the rules of thumb as Al Friedman says are meant to be broken. We tried to at least as the first set of experiments to look at common accepted values.

DR. SAUNDERS: As I understand, you ran sixteen experiments over a twelve months time period. Does that mean each one was made over a three-week time period?

DR. ORWIN: No, it does not. We initially, when we were in the experimental program, we would start it up and let it run a week and then take maybe the end of the second week we would start taking some measurements to evaluate some of our parameters to see where we were. And we found generally after two weeks we were at something we could call steady state, and we felt very comfortable about getting steady state. As we progressed we learned more about the system. We felt you know, we got to where we could say okay, we are going to change parameters to this and this is what is going to happen. We were still trying to show that objectively. We felt we found we could shorten the runs to almost two weeks, a week coming to what we called steady state and then essentially a week of experimentation. The systems we were using responded quite quickly and quite stabily, to changes.

DR. SUANDERS: You indicated in your initial slide, a rather dramatic change in microbial population, at least the color of the mass in one of your initial slides, between stages, would indicate a rather, at least, the potential of the most dynamic changes taking place over an extended time period with that unit. If we look at solids, like with activated sludge systems and that sort of key values that we usually use in evaluating those systems, I do not think that we would consider a seven to fourteen day time period long enough to turn a population over to get a stable result from the system. I have done similar units in similar situations in terms of nitrification with the units that did not nitrify and going to the units that did nitrify, it took something like a month to get that population in, you see the changes taking place. You see the changes in the nitrite, nitrate ammonia balances happening but it is an extended time period over which those change.

DR. ORWIN: We were seeing in that first stage the white gelatinous mass, we were seeing almost 95 to 98% carbon removal right there. Our second stage which was the light brown stage, seemed to be, I would say, a mixed, that was where the transition was taking place. The mixed bag - we would see sometimes predominant carbon removal and sometimes predominant ammonia, sometimes nitrification taking place depending on our conditions. But our system responded quite quickly. We were quite surprised. We were monitoring pH in the different stages. We were monitoring alkalinity in the different stages; we were monitoring ammonia and nitrogen, nitrates and others. We were doing COD analyses and they came right in and settled down. We were pleasantly surprised. It did not take a long time for transition. It may be a function of the type of feed we were feeding it, which was not a very complicated feed, a glucose, and we were not feeding it organic nitrogen; on might want to challenge us on that. Given what we were feeding it, it responded very quickly.

DR. SAUNDERS: Did I understand you to say that you pH was eight point five?

DR. ORWIN: We adjusted second stage pH to eight point five. We wanted to enhance nitrification.

DR. SAUNDERS: On what basis?

DR. ORWIN: Well, on the basis of the literature which was saying that was pH what eight and above was desirable.

MR. BUCKLEY: How did you measure film thickness?

DR. MOLOF: The film thickness was measured with a contrast microscope where we calibrated it and then just ground the depth, the focus, at a different depth.

MR. BUCKLEY: Any way of evaluating at the film thickness?

DR. MOLOF: What film?

MR. BUCKLEY: Active film thickness. Is there any way of evaluating active film thickness that you can think of?

DR. MOLOF: At this time the only thing that we could do, this was in the very early seventies, was to do the volatile solids, which as you know, does not really relate to the active. But what we feel is that film thickness and age show where the film does not become active anymore. After that initial 300 microns, it does not appear that the other material is doing the work.

MR. WATTS: I think I derived from what you said that in the operation of the unit, that you did not really change this speed or adjust for to get rid of the oxygen limitation. Is that right?

DR. MOLOF: We adjusted the disc speed from 10 to 40, but it is not in this paper.

MR. WATTS: Well, I mean are some of the things we are seeing as the load increases, systems below a certain level not being oxygen limited and systems above a certain level oxygen limiting takes place?

DR. MOLOF: That is exactly right.

MR. SETTLES: I may have missed it; did you identify what your mass loading rates were and how many COD per thousand square foot?

DR. MOLOF: Yes, I did, but I will be happy to repeat it. I said that the three grams per day which was what we were feeding for each stage, was about 15 pounds a COD per day per thousand square feet.

DR. REYNOLDS: Identified the DO limiting sections, substrate sections, did you measure dissolved oxygen and mixed liquor in the tanking sections that were oxygen limited, and if so, what were those values, what value did the dissolved oxygen have to reach before it was no longer oxygen limited?

DR. MOLOF: If the fellow who did the work was here he could answer your question. I do not...this was done in '71-'72 and he did not want to deliver it. So, I had to. We measured the DO but I do not remember the exact value. Incidentally, Dr. Chesner who spoke this morning has published data that shows a below about one point five to two. You will get an inhibition in reduction with discs, but that was separate work.

MR. STRATTA: I had a couple of questions. First of all, with respect to nitrifiers, did you try to enumerate any of the nitrifiers on your film?

DR. HOAG: No, we did not. We did not do any plating counts at all, and we really were not able to use them with photo micrograph or microscopic examination.

MR. STRATTA: And the second question I had, you made reference to having some problems in breaking up your film with blending techniques. What did you eventually use to break up your film so you could enumerate the organisms you saw?

DR. HOAG: Well, by scraping part of the bacterial film off into a slide, we could just separate it into various sections and on the slide itself, using a very large cover glass we were enabled to expose various edges of the sludge particles and there were some liquids associated with the biofilm that was on the disc as it was turning. And so we did get, we were able to examine the organisms that way.

MISS KINNER: What was the magnification of the plot on that shot you showed?

DR. HOAG: I think that was about 500 power.

MISS KINNER: Five hundred. And the other question I had for you is, you talked a little bit about the role of filaments and their increase in the second stage. What do you postulate is their, the reason for their dominance?

DR. HOAG: Well, I think the most obvious thing would be that they were preying on the bacteria that were oxidized in the carbon in the first stage.

DR. UNE: I would like to make a suggestion that you eliminate much of the bacterial part of your work when you do your thesis. First of all, you do not have many species and what you see in there are nitrifying bacteria N-debria in flat form, but the debria is not a nitrifier. It is a specific species of organism or genus organism and this is part of the problem that engineers have in effect bolt down on the field over the many years, and the microbiology of waste treatment. And I would like not to see it continued in the case of the RBC. I could not really get a good look at your photo micrography but I am very suspicious that you do not have Sphaerotilus or Nocardia in those pictures.

DR. HOAG: Well, as I warned, when I started, it is very difficult to identify the bacteria without plating them. You really cannot be sure by a microscopic examination.

DR. MOLOF: I wanted to ask you, did you notice any characteristics of RBC sludge if you look into a microscope and say the system is operating right?

DR. HOAG: I think I did, in especially in the first stage and second The second stages when we were really loading the system at a high carbon loading rate and a low hydraulic slow rate, there were really no other organisms other than the bacteria in the first stage, but when we decreased the loading rate or increased the flow, we would find a few species affiliates.

MR. PERRY: When you took the samples, did you try taking samples at different locations on the discs?

DR. HOAG: Yes, we did. The results that I was presenting here were all from similar locations on all the discs. However, a few times we did examine various sections of the biofilm. We found similar cultures but there were slight decreases in the microfauna population, as you went from the periphery of the discs to the center of the discs.

MR. PERRY: Did you try to quantitate the amounts that you took? You said you took it with a slide. Did you just reach in and scrape the biomass?

DR. HOAG: We really had a difficulty in, we were able to quantify the amounts that we took off the disc itself but since we really were not enumerating them, the numbers per hundred ml we really did not need to know what the amounts of the biofilm were rotating on.

Session 3. MUNICIPAL WASTEWATER TREATMENT
Presiding: Edward D. Smith
U.S. Army Construction Engineering Research Laboratory

DR. BANERJI: I understand that there was some problems in disc media.

MR. DUPONT: When I was there, there was one shaft that was out. The end of one had sheared off at I believe, it was at the drive unit. But it was laying sideways. All the other shafts in that path were turning so that it would not get lopsided but that had happened once before and also happened after I was there. So there is a definite problem. I think they went back and rewound it and replaced the shaft. I do not believe they planned on it because they had to tear the building roof off to go near it and replace it. So they had major problems. I do not know if they are having any more, but the data is dated. I wish I had more information but I am still going to school. But that is it, they did have problems and it was structural problems.

DR. SMITH: I am very sensitive to you but municipal plants; it is not only the RBC plants activated sludge, and trickling filter whatever, often times the operators only collect the data that they need to fill out their MPS permit and they do not collect the data that they need to evaluate the performance of their plant. And I think that is a very important consideration that the plant operators; what we should all try to encourage them to start doing that. Well the next paper stated that the water problem, water shortage by year 2000 is going to be, make the enery shortage look like a Sunday picnic. Now I do not know if you agree to that. Certainly in certain places in the United States that is true. Many islands have a lot of water problems and all of this type of research is in its infancy. I think it is very clear that one aspect of water conservation will be utilizing saline water for whatever, and I think this is a very important research. Also I think it is very interesting that we are using very sophisticated technology. For instance, she mentioned the electron microscope to study RBCs. And I think we need to continue in that area. And I know Dr. Poon who did the initial work on saline wastewater and RBCs has a lot of questions, I hope. Just one time I note that there is a lot of need for research on nitrifying bacteria and saline research in case you are looking for a topic NSF or EPA.

DR. SACK: I remember, I listened in at the early part of the session. What kind of salts did you add, could you tell me?

MISS KINNER: Yes, we added Utility Marine Mix artificial sea salt. We originally had thought of adding straight sea water but found we were having a problem with diluting out our BOD, diluting general organic and to avoid that we mixed up a concentrated salt solution with these utility salts.

DR. SACK: We have used high salt concentrations with coal gasification waste water and I wondered if some of the blobs on your screen, the dark ones, could have been precipitation when some of your added salts hit some phosphorous that you might also have added, or may not have added. Secondly, we have seen those cysts in there that you labeled as undescribed cysts and in the past I speculated they may actually be very stress protozoa or something of this nature that were getting ready to expire from the scene or had expire, and wondered if you found them present only under highly saline conditions and not under, not just the sewage.

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MISS KINNER: We found them present in both fresh and salt water conditions and Art Moore, I do not know if you are familiar with him, he is a protozoologist at the University, and he examined those cysts and he does not think they are stress for design. He just thinks that there is some kind of an insisting stage probably not even of a protozoan. He is thoroughly convinced that they are not protozoans at all.

DR. SACK: Does he think they are alive?

MISS KINNER: Yes, he thinks they are alive and we are not sure what forms they are but we are pretty sure that they are alive. And they seem to be present just randomly and that may be a function of the fact that we just missed them. Perhaps we missed it, but this is not a quantitative analysis and we have missed them. And the salinity that stresses as I say, they were apparent in both saline and non-saline conditions. As to the blobs on some of those pictures, I think also some react, it could be precipitate. Some of them comes from the fact that we did have a lot of material adhering to the disc that for the particular under certain conditions, and we think that that could be non-living material. Yes definitely, possibly precipitate, but we are pretty sure it is not living and in culturing we have been able to kind of eliminate some of those factors.

DR. SACK: It has been reported many times that high salts such as from industrial wastes cause effluence to become somewhat turbid. Did you notice that? If you did I did not hear you say that.

MISS KINNER: We did not have a problem with turbidity. That may be because the salt concentration that we were using were generally fairly low and we did put the salinity up to about six percent. We did find an increase in the turbidity of the waste and in fact the BOD was at that time, not up to the 30 milligrams per liter standard but quickly adjusted to that.

DR. SMITH: Did you think it was relatively high turbidity index for saline?

MISS KINNER: Yes. We were actually...if I showed you a fresh water and a saline sample, I will bet you could not even tell the difference.

DR. SMITH: I bet I could not either.

DR. POON: I have a lot of questions but I have only two general ones in salt expectations. First one, when, in most of them of course I use extra sea water and mix it with the sewage. And naturally it would have a lot of marine algae and sewage organisms on the biofilm. The disc biofilm has a lot of marine algae, some are vine colored, green color, it is very colorful. I regret because of the short time involved, I do not have time to study the nitrogen removal and phosphate removal. And I would recommend anyone who would do this, we should look into that. Maybe there is additional nitrogen and phosphate removals better than using fresh water. I do not know if you have data on the quantitative basis at what concentrations they may be, I should say the fluctuation of the concentration would have any effect on the growth of various organisms because this is the only thing that I found within the range from zero, I should not say zero but a hundred parts per million up to about 12,000. If I ran the system long enough to the steady state, there would be no effect at all. But if I let the concentration fluctuate, the hydraulic loading will highly affect the BOD removal.

MISS KINNER: We do have some data and we did not find any inhibitory effect with changing concentration, but of course these things were brough up on changing concentrations since we changed the quantity. It was constantly varying between 2 percent and 4 percent. So we did not acclimate them at all to a straight salinity as part of the work. They are constantly in a varying salinity. But we did not observe any inability of the microorganism to treat the waste under those variations as I explained in the last question. We did see some limited ability to meet on an initial period of acclimation and then we did after that 6 percent or 5 percent.

DR. POON: Where were these experiments of concentration we have looked after?

MISS KINNER: We did have percent solids at all on that, so we just measured the influent and effluent salinity. That is what we were monitoring and it was pretty much between 10 and 12 points per thousand.

DR. POON: And your soluble BOD's were around 15 in effluent?

MISS KINNER: Yes 10 to 15.

DR. POON: And you may or may not meet 5 mg/l effluent requirement in BOD totals?

MISS KINNER: Well we did do totals as a matter of fact but I just did not mention in here. They were all under the 30 milligrams per liter. I just drew a purpose for this study. I would also say that besides the bacteria we had, the protozo, we were pretty sure that those were not in green form present. We did not get the algae at all. So this could have been there in limited numbers. But anyway, we are using a salt water waste. The higher organic concentration should knock off the bacteria and this has also been seen by marine microbiologist Dr. Johnson at University of New Hampshire.

MR. VESIO: At the point where you were testing, you switched over to a saline solution where you have a salt water intrusion. Was there a drop off in the overall treatment effect until that point, or if there was, how long did it take for the bacteria to acclimate themselves and bring the treatment level back to what it was initially before salt water intrusion?

MISS KINNER: Well first of all let me say that we initially started the experiment, we used a COD as a monitor, a daily monitor, instead of BOD so that we can have lag time. And this COD did not work at all in salinity, so that the first experiment that we had, we actually assumed that after one week we were at a steady state and found out that were the BOD that we were in fact. That steady state we were getting ridiculously high answers to the COD, if you had not promised to run one of those, we should have had similar results. I spoke to a gentleman who said they actually ran TOC which we are now switching to. So it is hard for us to say how long it really took to get the steady state. The second one that we switched to the salt. We did find that we got the steady state faster than a week, in fact, we were getting good BOD removal in four days or so, but we just do not have the data because of these testing problems that we ran into on solids as to how long things have to acclimate. From a microbiological viewpoint, just salt sampling that I was doing, there did not appear to be any specific change in the populations but we were observing as a die off of rotifer and protozoa.

MR.VESIO: Does 6-7% salt concentration significantly affect bio-activity in RBC?

MISS KINNER: Yes, we did that at seven percent. There were variations because of the initial problems in running the system. We had some variations that we kind of were fitting into there up to six percent salinity. And the system did take a few days corrections, kind of stepped to a quota effluent and get most of the BOD did handle that way. But the removals were somewhat higher, excuse me, the removals were somewhat lower. The BOD were somewhat higher during that period, and they probably would not meet and actually load for this municipality.

MR. FREEMIRE: Is there any subsequent follow up on performance as to adding up the alkalinity to adjust pH or aerating?

MR. HITDLEBAUGH: As a result of the work that we did, the Army Corps of Engineers both from our district had put in a project to add supplemental aeration to the system and that was, that is supposed to be under way now but I have not checked on it recently, and hopefully that will be in next summer. I am not sure what the status is as far as any pH adjustment with the system. They may have been waiting to see what the effect of the supplemental air had to see how much of an effect that would have as far as improving all systems.

MR. BRACEWELL: I did not quite understand the impact of temperature. You said that at 13 degree centigrade that you were having an effect on your removals. The data during the winter time was about as good or better from the summer. And so how do you know, what makes you say the temperature was important and how important was it?

MR. HITDLEBAUGH: Maybe that was more confusing. What I was trying to say was that during the summer time the temperature is 26 degrees centigrade, much more optimum conditions for biological activity. During the winter time, the temperature was 13 degrees centigrade and much more an inhibitive factor on biological activity. Even now, I am saying a negative effect, from summer to winter even though the temperature has an apparent negative effect because of the fact that we did not seem to have any problems with the DO limiting conditions, for some seemed to be better. So what I am trying to say is that the fact that there was more dissolved oxygen limiting conditions in the winter time, over compensates for any negative effect due to a lower rate of temperature.

MR. BRACEWELL: Did you see any interim effects, for example in the fall when maybe the temperature was 2 degrees centigrade?

MR. HITDLEBAUGH: Yes, definitely.

MR. BRACEWELL: Was not a DO limited?

MR. HITDLEBAUGH: Yes, definitely.

MR. BRACEWELL: Is that a BOD removal?

MR. HITDLEBAUGH: The plant did meet and I mentioned the word periodic, we were involved in the process from August through January and in fact Autotrol even went down and did some, a little bit of testing and in fact during their test, the plant was needed experiment. So there was a few periodic problems and we picked, we chose the study to look at the peak summer and the peak winter conditions to see the extremes there.

MR. SHAMITKO: We have received some reports that some of the shafts were breaking with some of the Autotrol biodiscs. That is my first question. Has that been corrected and if it has been corrected, what was the original problem? And my second question deals with your samplings for the BOD testing. Was the sample taking directly after the waste had flowed through the disc or after it went through the final clarifier?

MR. HYNEK: You were speaking of my impression of the talk.

MR. SHAMITKO: Either gentlemen.

MR. HYNEK: The constant situation we were reporting to you raw wastewater and final effluent because that is what was reported by, according to the State of Michigan. In the Cadillac situation we had information at both points, the secondary clarifier effluent, the biosurf effluent, and the final filter effluent for ammonia. Did I answer your question?

MR. SHAMITKO: Yes sir.

MR. HYNEK: In the case of Cadillac it was flow composited sampling automatically. Very pretty plant. We all had shaft difficulties and we have faced the steel failing, the construction of a shaft requires a welding procedure with a backing strip when you bend two U-shape metals together. And the failure was placed to an interruption of this backing strip and not according to specifications by the way, and we have since terminated our arrangement with that particular supplier. We are currently purchasing our shafts from steel-making people in Japan and with a very superior shaft, and there is a definite program with Autotrol Corporation and its customers. But those that have failed are being replaced. It is not necessarily all at Autotrol expense but it is a very, as you can imagine, a very serious situation when one of these large shafts break and causes the customer concern. But they have been very tolerant and we have been open as we can with them to explain what happened, why it happened, and what we are prepared to do to back up our product.

MR. SHAMITKO: One other question. The follow-up of the first answer you gave. The BOD removal rate, was there a big difference from the BOD measure after went directly to the RBC as opposed to after it went to the clarifier? Was there a big difference between the BOD amounts?

MR. HYNEK: Well, we do not normally get separate data like that from the customer but the thing is how you take your sample at the biosurf effluent. If you do not settle it you will get quite a large number indicative of the endogenous respiratious. What we do usually at that point is to settle it for 30 minutes and decent the supernatant & clarification and then compare that number to an actual clarifier performance. For example, they are adding alum for phosphorus removal. The clarifier is going to do pretty good. And our set of samples probably would not settle or give a BOD number as low because of the little bit more of suspended solids contribution.

MR. GARY: I think the area of the small package treatment plants including RBC. In the state of Pennsylvania, the state may accept your treatment system if the system is approved by NSF. Is your company planning to get the NSF to approve your equipment?

MR. VESIO: Well we certainly thought about it. We have not done or pursued NSF approval with this product as yet. We have mixed emotions about it. The NSF approval even though we realize that many states do require a pre-engineered type of system, what we like to think is the fact that number one, in the RBC approved of themselves in terms of performance capability. Number two, we certainly would be the first to admit that there is nothing magical or mystical about the concept that we have come up with. We are fortunate in that we are able to take the puristream technology with regard to the steel tankage and diffuse their portion of their system, as well as Bob Joest and the TAIT Bioshafts expertise in terms of the size in the RBC, and take the two technologies together and that is a good system. If we find that because of that is certainly a pre-engineered type of package. If we find that NSF testing of some type would definitely be required, quite frankly we would have no recourse but to go along with it because we have a tremendous amount of confidence in this particular product in the concept.

MR. CROUCH: What kind of mechanical collector do you have in that final clarifier?

MR. VESIO: On most of our models we use a hopper bottom type clarifier such as the typical arrangement show. When we get up to above 150,000 gallons per day, you have the option of going with a rectangular suction type of collector mechanism or a rectangular chain flight collector. But those are often standard and flows from 150,000 gallons per day and up.

DR. POON: I wish you could elaborate how do you size RBC system in the final clarifier?

MR. VESIO: The RBC themselves, we feel, that we have a pretty conservative design in that in the first stage reactors, we do use a hydraulic loading rate for domestic strength sewage, and in the first stage units, the hydraulic loading rate ranges from 1.76 to 2.0 gallons per square foot per day. Now as far as the clarifiers are concerned, as indicated the surface settling rate does vary depending upon the flow. What we were striving to do was to supply the clarifiers with two hours retention time. Now some states have their own criteria for clarifiers and my feeling in this particular thing since I had input to this particular concept was that the two hours is plenty really even in an activated sludge system, two hours is plenty of retention time with any clarifier. With a RBC system, in most instances especially if it is a lightly loaded system which just about our model upon start-up and quite frankly unless something unusual happens in the development of the community, these systems do not usually come out to full loads for some length of time. And because of this the ambiant temperature is going to be given nitrifiers and in the tail end of the third stage and in the first stage and a first stage reactor. You are going to run into a problem if you get into an extended retention time in the final clarifier where denitrification is going to begin to occur. The Autotrol takes that into consideration; the post-aeration aspect problems is great and if you keep in mind that we are talking applications involving recreation parks, subdivisions, and so forth. But it is only in which the problems are apparent in the inheritance of the extended package plants. So basically in summary, is that in our large first stage reactors we do not exceed two gallons per square foot per day in hydraulic loading and because the clarifier is a standard size to give you two hours retention in both the primary and the final clarifier. If the state regulatory agencies or such that no retention time is required, you have the flexibility to do this because of the marginal configuration. This is true for the flow equalization chamber, it is true for the aerobic digestion chamber. As I indicated the two percent on flow equalization to keep the separate capital in aerobic digestion. There are states like Ohio, states like Indiana where surge capacities in Indiana, in a school project they require 67 percent surge capacity at the time of school. With this particular margin of configuration, you can give them 67 percent. So Ohio has a very elaborate program that you go through, based on an average daily flow and a 3.3 peaking factor and you go through the calculation that they prescribe, and you could come up with a flow equalization chamber of about 37 to 40 percent in volume. So you can change the volumes as needed to meet the state requirements or application requirements.

DR. SMITH: We would just like to take 30 seconds to sort of summarize what we found from this conference. I think that RBC technology is demonstrated as that to be an advanced technology and indicated by the number of manufacturers and competitiveness among these and this sort of gives us a choice. Also you notice there seems to be a real wide interest in RBC and there seems to be so many applications for RBC that most other technologies do not have that many applications, lechate treatment, nitrogen and phosphorus removal. I just returned from talking to the puricycle people in Colorado and they had a very interesting scheme where like homes in the mountains where you do not have a well and you cannot have a septic tank, but they have come up with a treatment channel with a little RBC unit in it and some very rather sophisticated technologies. And it is monitored with by method of process or technology and it costs about $8,000. That is actually cheaper they claim than buying a $30,000 small package treatment plant. It takes their sewage and treats it to portable water quality. And I am certainly not an oponent of portability at this time. There is a lot more research that must be done, but I think the little black box that we have heard so much about is finally here and in that NSF is evaluating it at this time. Also you learned from some of the papers that there is a lot of clearance in modifying some of this RBC technology and it is nice that we have RBC technologies of all different sizes for households or trailer park or service stations on up to hundred million gallons per day plant.

Session 4. BIOKINETIC STUDIES
Presiding: C. P. L. Grady, Jr.
Department of Civil Engineering
Purdue University

DR. JENKINS: Mike, I really do not know how to phrase this question but the impression I get is that you are forcing the biodisc system to be like an activated sludge plant by intentionally scraping off materials from the surface of the biodisc and turning it away just as you would waste activated sludge, but I guess my general question is how does this, how does the system that you describe, this method of sludge wasting really replicate what goes on in a biodisc system in the field where this type of wastage does not take place. And I guess the specific part of the question would be, how would you relate your sludge wasting rates conducted as you did, to the nitrification values you obtained with them, to what you would get in a field scale system where hydraulic loading or organic loading seems to be the parameter, used to correlate with whether you get ammonia or not, or oxidation or not.

DR. SAUNDERS: I do not in any way mean to indicate that I would like to have the operator go out and scrape his disc system. That is not the purpose, that was not the intention. The purpose was to measure the rate of growth in the system and to see of the growth rate as we would relate it to SRT values, or any other way we like, relate it to the performance of the system with respect to nitrification, so we are not talking about this as a means of operating the system or in any way applying it to practice. I do not want that to be indicated. The other side of it is, it says simply the data that we have for nitrifiers can be applied to these systems. And that was really our purpose, to simply see if the relationships that we developed forcing the systems would in fact indicate that the kinetic data that we reliably used in activated sludge systems could be applied here. There is a tremendous amount of information that has got to be combined with this. We have got to look at the heteretrophic organisms and the rate at which they grow in the systems, the rate to which they are being sloughed from the system, there is a tremendous amount of work that needs to be done. Probably one of the most important things that needs to be done in this regard is to look at the dissolved oxygen, because now we have been looked at probably the primary reason for lack of nitrification, or having nitrification in the plug flow system, the first one is overloaded to produce anaerobic conditions. If we do not have dissolved oxygen we do not get nitrification until the subsequent unit. The results that we have got, again, if we listen all of them this afternoon, they have used this approach and my contention is it is a valid approach, to look at kinetic growth relationship within the culture system. You have got to combine it with other relationships, with the growth of heterotroph, experimental determine and also relationships for oxygen uptake and oxygen profiles within the film.

DR. POON: I am not too sure about your last statement. If we ignore the consideration of suspended solids effects on safety because such a high solids exist in the first stage of the RBC, oxygen demand would be very high. That could probably, easily reaching a critical factor in the design. That is just one comment.

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DR. KINCANNON: I think you are very true on that end. Again, at least my feeling is we should be looking at the suspended solids and again I made the comment that, well, maybe a factor of safety force, but like you say, it could be a big problem too.

DR. POON: I want to ask you, did you look into the sludge retention time in the system, as a whole, and then try to differentiate between two fixed and suspended biofilm?

DR. KINCANNON: Well, I tried to omit the problem that I had in trying to make the calculations. Now, again, I take the different approach than what Mike did in the last paper. He takes the total solids. I am interested in only the solids that I considered being produced. I look at sludge retention as being the reciprocal of a growth rate and not the time the sludge is held in a unit. You know, it is a growth rate, when you have this solid accumulation such as this, it is difficult to determine the actual amount of growth rate. I tried to make some calculations on this and ran into a lot of problems. Now, I agree very much with what Bill Characklis mentioned earlier, that you finally reach an equilibrium level so that your growth is actually constantly being lost, and if you, and I do this with a biological solids all the time, you do not have to worry about this accumulation, but the suspended solids in the effluent basically measures what is being produced that day.

DR. POON: That would be a constant sloughing off on the biofilm sometimes in fact some of the particulate form would be trapped by the biofilm. There is a constant change between sloughing and attachment to the biofilm. Well, my calculation for sludge retention time is close to 20 days under perhaps a similar loading as yours. I just want to see if you have any calculations concerning it.

DR. KINCANNON: Well, no, well, I made them on this but I was, like I say, I was concerned about the numbers because of the solids being retained here.

DR. REYNOLDS: You mentioned an active biomass thickness, I think it was 200 microns. How did you determine that?

DR. KINCANNON: From the literature. I did not determine this on our system itself.

DR. DENNIS: When you calculated your growth rates you subtracted the delta COD from the effluence? I believe that was your technique, you got the refractory compounds by a delta COD technique.

DR. KINCANNON: This was for the substrate removal or SB. SB was the total COD minus this residual COD. To me this more nearly represents the amount of material that is available for biological degradation. In other words, we have this residual which, if you let it aerate for a week you are not able to lower it below that.

MR. DENNIS: So, you use the delta COD to calculate the effluence, and that was what you used to get the unit of substrate utilization rate?

DR. KINCANNON: Yes.

MR. DENNIS: I do not think you subtracted it from your influence concentration, or did you?

DR. KINCANNON: I did not.

MR. DENNIS: Would not that tend to bias your results to a higher substrate utilization by subtracting it?

DR. KINCANNON: Well, there is a question of whether you should subtract it from the influent. There we were using glucose as the substrate and assuming this is completely bio-degradable, that is the reason I did not.

MR. KIM: What I am asking is if you include R in your equation do you think that your model becomes more useful?

DR. HAUNG: Well, I do not really know.

MR. KIM: Actually, this film thickness is proportional to the tip velocity rather than rotational velocity. Why do you use the rotational speed instead of the tip velocity?

DR. HAUNG: Well, I have N and R time together. Of course, it is also a function of kinetic speed too. I did not neglect that.

MR. KIM: If you use tip velocity maybe you can generalize your model.

DR. HAUNG: Well, if I use different disc sizes I may do that. In this study I only have used one single size. As Dr. Friedman stated last month that the rotating speed should be considered at the same time for scale-up. That is very important. You do not want to talk only tip speed it should include both the radius of the disc and rotating speed.

MR. KIM: In Dr. Friedman's paper last Novemeber he mentioned that the rotational speed was a very important function in his research, right? However, your model is really related to the film thickness. The thickness is a function of tip velocity, rather disc speed. But that is not my point.

DR. O'SHAUGHNESSY: What was the minimum ammonia effluent concentration you achieved?

DR. HAUNG: You mean the whole test?

DR. O'SHAUGHNESSY: In your effluent?

DR. HAUNG: The minimum?

DR. O'SHAUGHNESSY: Yes, what was the lowest ammonia concentration?

DR. HAUNG: In the effluent the ammonia concentration is about one mg/l.

DR. O'SHAUGHNESSY: Did you go below one? Most of the time you were above two?

DR. HAUNG: Yes, most of the time.

DR. O'SHAUGHNESSY: Let us go back to the previous question. In other words, we have done fixed-film reactors also. To get zero on the kinetics above 2 milligrams per liter, but once you go down below two or three it goes to first-order kinetics.

DR. HAUNG: Yes, that is a possibility.

DR. O'SHAUGHNESSY: And looking at the way you had it, I think the model would be valid for your higher concentrations but to try to extrapolate that to get that final removal might be dangerous.

DR. HAUNG: I am not saying your comments are wrong, but some further studies should be conducted to reach the possible conclusion.

DR. O'SHAUGHNESSY: What was your minimum influent concentration?

DR. HAUNG: Minimum concentration is about 3 or 4 mg/l.

DR. O'SHAUGHNESSY: Your minimum influent concentration was 5 in the model held as compared to only a hundred milligrams per liter.

DR. HAUNG: No, I did not use both concentrations to fit that small unit. I did not try that. I used a very high concentration in the full scale study up to 600 mg/l.

Session 5. AIR DRIVE AND SUPPLEMENTAL AIR
Presiding: W. A. Sack
Department of Civil Engineering
West Virginia University

MR. GERHARD: I noticed in the beginning of your talk that you indicated that there were two methods of control. One was speed and the other was aeration. You have given a history of some eighteen to twenty four months with work on supplemental aeration, but you did not mention at all the speed control. Did you, is there a reason for that?

MR. SULLIVAN: Yes sir. We feel from a standpoint of power evaluation, believe that RBC an increase in speed increases power exponentially. We do not feel it is feasible to increase the dissolved oxygen within the reactor by increasing speed because the power requirement due to an exponential rate on the BOD removal film

MR. GERHARD: The contrary situation might exist, though, because you cannot get air for nothing either and in terms of energy you have to provide blowers certainly with the sufficient size. As I understand Alexandria situation was a mechanical drive and you added air to it, which improved the plant efficiency certainly at the expense of the air blowers. Can you positively state without having pursued it in depth as you have your aerosurf, that the balance of energy is in favor of air as opposed to perhaps a minor change in the rotating RPM with a minor change in horsepower?

MR. SULLIVAN: Based on our conclusions of work with air and at one point six RPM, determining the power evaluations of increased speed, it is my conclusion that at this time that we can definitely state that we definitely do not want to increase speed on RBCs.

MR. GERHARD: Is there any kind of perhaps kind of a mechanical reason that you do not want to exceed a sixty foot per minute tip speed or whatever?

MR. SULLIVAN: The basis for my statement is based on pure energy operating economics.

MR. BURNER: In your 18-month testing at South Shore did you look at the differential and settleability characteristics of the sludges?

MR. SULLIVAN: There was during, I do not want to say every test stage, but during the majority of test days discrete particle settling test done on the effluent from fourth stage. The results indicated that the settleability of overflow rate reach an effluent concentration or approximately the same. In some cases the mechanical drive did performance and in other cases the air drive did out the performance but the ranges of overflow rate required to get a given effluent concentration at the same rate they did it by approximately fifty gallons per day per square foot, which I feel is the accuracy of the test.

MR. WONG: You have not mentioned in your discussion regarding the amount of air required for supplemental aeration

MR. SULLIVAN: The results here do not present all the data evaluated. There is another paper called the aerosurf process that does go into all the data but I could tell you that the range of applications are there on the data that was presented, was approximately a hundred to two hundred CFM per shaft at the low loading of approximately a hundred CFM to two hundred CFM was only present on upstream stages generally downstream. We were in the range of eight CFM, but an average of first stage of approximately 140 CFM, 150 and downstream 80 to 120. It is very difficult to give that an overview. What you have to do is to examine nine months data to find out what the organic loading are to find how to match your organic profile with the amount of influent flow but the data is available.

MR. PATRICK: How do you arrive at biomass thickness?

MR. SULLIVAN: All the biomass thickness that was conducted at South Shore was conducted by rotor shaft bearing. In one of the system, was jacked up you know the dry weight of the media with biofilm on it immersed with a 40% immersion. We then weigh the bearing and we get a reaction weight and that tells us the total slime on the media.

MR. PATRICK: And how do you equate that to a thickness?

MR. SULLIVAN: Well you have x amount of square foot on it, the media, as present and each cubic foot of biology was approximately, weighs approximately 62.4 pounds of weight. You have a hundred thousand square foot. You know the amount of biomass present has a weight of "x" multiply a hundred thousand square foot times the thickness to come up with that weight, it comes up with the biofilm thickness.

MR. PATRICK: And is there some advantage to minimizing biomass thickness other than operational considerations?

MR. SULLIVAN: We have definitely found in the minimization of the anaerobic layer tends to promote a biology which was higher in life and has a better BOD removal in the aerobic layer.

MR. PATRICK: How about maintaining the structural integrity of the system?

MR. SULLIVAN: Well, there is no question you do reduce slug you reduce structural integrity, so there is no question about that.

MR. PATRICK: Do you have a cutoff weight on that?

MR. SULLIVAN: Yes, thirty thousandths.

MR. PATRICK: What was that?

MR. SULLIVAN: The test data that the Autotrol has performed were being an operating shaft life of fifty years of biomass thickness of ninety thousands of an inch. We have conducted all our structural requirements at ninety thousandths. For a process standpoint I think the number I quoted was sixty thousandths, because ninety thousandths biofilm thickness is too thick.

MR. HALLHAGEN: How does the temperature affect the supplemental air system?

MR. SULLIVAN: The work here was presented independent of temperature. It is obvious that warmer wastewater temperatures and warmer the air temperatures the amount of air will increase. The data here was over an eighteen month scan so as a result it did take in winter and summer operations, but we did not go through a number of seasonal changes, so I really have not determined the effects of wastewater temperature on the amount of air put into the reactor. It would take probably another year to do that. There is no question in my mind however, that a lower wastewater temperature the less air would be required and at higher wastewater temperatures more air would be required.

MR. RUSHBROOK: Does Autotrol see itself moving away from mechanical drive units in favor of the air-drive?

MR. SULLIVAN: There is no question from the process standpoint that automated feels that aeroreactor, is the right way to go from the biological standpoint. I think all our recommendations are that way and have been that way for apporximately six months to a year. We are recommending in certain installations retro-fitting of existing installations with aeroreactors. We believe it has definite process benefits. Yes, we are going on it.

MR. MOORE: I have two questions. One question is with the air-drive. What is your recommendation of maximum size with BOD loading compared to mechanical? The other question is, it seems to me that as you get into second, third and fourth stages of the process that you are not dealing with the biofilm thickness problem and thereby it seems that a need for air-drive is minimized.

MR. SULLIVAN: The current design with the recommendations are approximately five pounds soluble BOD, with the air and four pounds soluble BOD maximum stage level with the mechanical. You can see that the data presented indicates about six pounds before you get a DO sag. But we design it right now on a five pound, basically because of the data we do not have a temperature, extremely high race with the temperatures. Well in essence if the biofilm is thin the effects of air are minimized and you can see the reduction. There is not a significant increase of BOD removal. Autotrol feels at this time that energy is a big function of the RBC system. The capability to incorporate air and to reduce rotational speed, and to minimize energy on the downstream stages is critical in the future. Air drive gives us that opportunity. From a process standpoint, downstream at low loads, we do not see a definite need for process improvement. However, coupled with the reduction of speed and the optimization of energy, we feel that air is critical.

DR. CHESNER: Is there any real concentration of dissolved oxygen that should be maintained in the reactor?

MR. SULLIVAN: Very difficult question to answer. Off the top I would say about a part of a half a part of dissolved oxygen within the fluid. The critical point is really dependent upon the biology present on the media. When we see a biological shift we try to raise dissolved oxygen if it is a half a point or a part. In other words, when a biology shifts from a truly aerobic culture to an organism, it is like Beggiotoa or something. It is oxidizing sulphur deriving its energy sources from that other than carbon. We tend to change the environment and whether we change the environment the only control that we have is increasing the dissolved oxygen within the reactor. So the dissolved oxygen is not the critical criteria. As an off the cuff statement I would say about a part to a part and a half, but the real key function is the biological organisms present the change in that organism.

JR. JANK: Just a continuation on that question in Figure 2 you showed the DO level with the...applied organic loading. In order to get one milligram per liter of DO you showed a loading of one pound.

MR. SULLIVAN: No, it is a delta in dissolved oxygen for instance if I was entering the reactor it begins to sag dissolved oxygen. If I was entering the reactor two it determines when the biology utilize oxygen. If I was entering the reactor two when I said I would have a negative. If I was going up I would have a positive. What I am finding out is when a mechanical driven system I am sagging in about three pounds soluble and on the air driven systems I am still increasing dissolved oxygen at approximately six pounds soluble at an increase in the RPM in the reactor up to a six point soluble. It is not concentration, it is delta concentrations.

MR. BEISEL: In your studies there is a cost for the supplemental air both the capital blowers and piping plus your operating costs. Was this cost effective per pound of BOD removed, in other words the additional cost, if I wanted to convert my present system?

MR. SULLIVAN: Well, I guess the question is on necessity. I think on the newest systems there is no question on mine with the potential of energy reduction and downstream speed control and flexibility. We feel it is a necessary capability in an RBC system with existing plants. It depends on how the effluent is doing. We have existing plants that we definitely are recommending in order to achieve our full quality to incorporate supplemental air. If you are currently having no problems with the mechanical equipment there is an awful lot of plants mechanically around the country now, right now, doing extremely well. I would not put in there, but if something happens that you are unknown of I think the incorporation of air could provide a significant benefit to the reduction kinetics. I would not recommend it to take every mechanical drive right now to combine with air. I do not think that that is necessary.

MR. BEISEL: I have a 3½ MGD plant with 24 Autotrol units and I have had some operational problems caused by other things in the plant, such as high strength supernatant from the thickener that is giving me some problems. I have finally gotten around it but I have all mechanical drives, have no way to supplement air without additional blowers and piping.

MR. SULLIVAN: The first step that Autotrol recommend is at this particular time, to optimize the system as much as possible with the mechanical. In the event that cannot be done the incorporation of air is probably a desirable end result. But I do not want to tell you at this time that you should go out and spend "x" number of dollars on this.

MR. BEISEL: Well, I only have about four months operating experience on the plant. It started up in October.

MR. SULLIVAN: Maybe we could discuss it in detail afterwards.

MR. BROWN: Did you evaluate how much air it took to ratate the given speed or power requirements?

DR. SRINIVASARAGHAVEN: We certainly did not do anything about power requirements because the air was provided by a portable compressor so it would not give very much indication. In terms of air requirements, it escapes my mind now but I think the one point six RPM took 180 CFM. At one RPM it was close to 120 or 130 CFM.

MR. BROWN: Will air be the sole source of power to drive these or will there be any form of mechanical assist?

DR. SRINIVASARGHAVEN: We have not provided any other source. We are envisioning a mechanical device that would be serving each train of RBC that you could hook up in case you have problems to start the RBC. But each RBC shaft will not be provided with a separate motor of some sort.

MR. WARD: Was that an overall loading of one point four pounds of soluble BOD per thousand square feet?

DR. SRINIVASARGHAVEN: That is correct. The first stage loading would be four times as much.

MR. WARD: I thought you showed six stages though, instead of four.

DR. SRINIVASARGHAVEN: But in the pilot unit I am talking about the loading was four times as much in the actual installation. You could operate it anyway you want. You could lift the baffle, you do not intend to operate six stages. You could operate any number of stages from zero, one to six, so the loading would depend on how you wish to operate, but you would not intend to operate any higher than five pounds. First stage loading.

MR. WARD: I missed it if you presented it, what is your general conclusions about reducing the RPM of the latter stages?

DR. SRINIVASARGHAVE: The pilot unit was not set up to reduce the RPM for each stage. You could reduce the speed only if all stages...the one RPM test indicated test data, indicated performance is just as good as one point six RPM. So unless there is difficulty in meeting the effluent standard we will recommend that they operate at one RPM with the provision it is fairly simple to increase the RPM, so they could increase it.

MR. SERPA: On your soluble BOD was this a calculated value or an estimated value based on the total BOD?

DR. SRINIVASARGHAVEN: It is a test value. We measured the soluble BOD and projected for the future the design values projected at the test program values it actually measured.

MISS ROSENBERRY: I have two questions. The first, if you put the RBC over the activated sludge tank, is there any problem in getting at the air header if you have to?

MR. COWEE: I would say that depends a great deal upon the type of air header that you have in the system. I am not really familiar with activated sludge system so I cannot specifically answer your question. I would say there is a good possibility under some circumstances that could be the case.

MISS ROSENBERRY: And the next question is, is there any change in head by adding the RBC?

MR. COWEE: Not significantly.

MR. GERHARD: I have got a couple of questions on the Philadelphia thing, having listened to the talks several years ago at the Philadelphia WPCF Convention. From the claims made at that time, seem now to have been either changed, altered, corrected or something as I understand it, the fine bubble diffusers were put into the Philadelphia twenty-two shaft RBC installations to bring the process up to the required design improvement level. I heard from another consulting engineer who was interested in the same or a similar process that the chances were good had it been checked, that just the addition of the fine bubble of diffusers system alone would have given the improved capacity or treatment levels, and that the RBC air drive units were superfluous. That is the first question. To what extent then do you know or have you compared the improved process performance without the RBC is running. That is number one. Number two, I understand also that the same installation is preparing to take bids on 180 more shaft with mechanical drive and would you care to comment on that?

MR. COWEE: I think I will take the last question first. I am not really capable of answering that question because since it is now a system that is coming up for bids it is being handled through the marketing department. I had not been related with the Philadelphia system I would say now for a good five, six months. I do not know what is going on as far as hanky panky related to bidding the system at this point. The only indication that I have or that I have heard of it at this point, was that due to the requirements of maintaining a very specific RPM profile on the air drive, the operations personnel were not interested in continuing to do that type of fine tuning on the system, but all the testing that we have done and that we have seen done by Philadelphia has indicated that there is no need to continue doing that fine tuning. So it is more or less a political question at this point, between the operation people as to whether or not that is the case. That is my personal feeling. I really cannot say specifically what is going on there, so I cannot answer your question directly. I wish I could remember all three of your other first question.

MR. GERHARD: Was there a comparison of the RBC not in operation with the fine bubble? Was there an actual comparison made?

MR. COWEE: You mean having the RBC in the tank without turning them and using the fine bubble, the RBCs. will not sit in the tank without turning. Just the oxygen pickup from the air pickup from the fine bubble system or the course bubble system and the hydraulic role does impart motion to the RBC. So they have to be tied down in some fashion so as not to rotate. I do not think that has been done, I am not aware of it at this point.

MR. GERHARD: How about without the supplemental air condition? Was that turned off to see what the natural effectiveness?

MR. COWEE: I have been told that the people at Philadelphia had run the RBC system in the aeration tanks without using the supplemental air header and were able to get significant rotation of speeds on the shafts and were able to achieve treatment on the shafts that way. Once again, I have not direct knowledge of that. I was not involved in that testing.

MR. GERHARD: You indicated that you were using about 3½ out of the total available horsepower of 7½ which would leave 12 horsepower extra motor capacity, that you then used to drive blowers.

MR. REH: That is correct. Yes sir.

MR. GERHARD: Was any attempt made at least on one or two units to increase the RPM by simply changing sheaves or sprockets on a chain whichever way you wanted to do it, utilizing a full 7½ horsepower than in rotational speed to see what effect it would have. And as a side comment on that, we have done so on pilot and full scale units, we have found that the biofilm thickness is very definitely controlled by RPM because of the additional shearing forces and stresses of the higher speeds. We found additional aeration, we found additional sludge settling characteristics change, in other words better suspended solids and so forth. It seems incredible that you did not at least try this. And I tried to ask the gentleman this morning the same question and it seemed to be an evasive answer. You had an opportunity to do so, was it done, and what were the results?

MR. REH: I will not evade your answer. The direct answer to that is no, we did not try. It has not been excluded, the discs were there and will be there until project engineers come home. This was an experimental program, only so many things can be done over a given period of time. We did discuss that option with Autotrol and it was felt that with what we could get in terms of increased revolution with what we had to turn it with, we would be better in the short run to go with what we did. There is no reason that we cannot go back and try it and we probably will.

MR. PRICE: It seems that you really did not have a choice. You give a cost analysis to show that supplemental aeration is the cheapest way to go, but with biofilm thicknesses as thick as they were, if you did not control the biofilm thickness it seems like you were looking for impending shaft failures.

MR. REH: THat is a good point, and the Autotrol folks again were helpful in making some load cell weight measurements and the equipment they were using was some prototype stuff and they are not quite satisfied with all the numbers but they did indicate that if we kept up with the thickness of the growth we would probably be increasing our potentials for shaft failures, so yes, we may be trying to justify the obvious, that is a good point.

MR. PRICE: The other question is did you do any DO profiles, that is was the improvement due to higher DO or was the improvement due to the shearing process of the aeration.

MR. REH: Two fold. This system in my opinion was oxygen limited from the word "go". We found the oxygen measurements before aeration, Roger, generally were in the range of one to two tenths on the influent side and in no case did they exceed a half a part per million or so on the effluent side. When we begin to aerate we found we consistently found that we consistently maintained a dissolved oxygen in excess of two and sometimes as high as four and five parts per million, so we receive a benefit from the actual aeration of the system as well as from the turbulent supply. It is two-fold benfit.

MR. JOHNSON: I understand from what you said there, that the baffles made no difference. Are you saying that you tried baffles between each shaft and without, and you are saying there was sort of a plug flow effect in either case?

MR. REH: Correct. The baffles, these were between each of the four shafts in the series we have a set of five removable baffles plates and we tried every conceivable configuration from all baffles in place, to no baffles in place and the results there indicated less than two parts per million soluble BOD. And I think plus or minus two parts is questionable in that sort of a test anyway.

MR. GERHARD: Again, I have some comments, Jim. First of all, we conducted pilot plan studies of a similar nature and experienced similar results, so we corroborate your findings on your Cold Spring thing. Secondly, we also found that by not really heard it described as a closed media as such, that before but none the less there is a definite difference in water passages and air passages in the different manufacturers of plastic designs. We have likewise not experienced problems with the oxygen limitation on our units and where we have had this situation we have had no problems with meeting performance requirements and so forth. Do you know or have you checked out or do you know of any place where a study is planned where a comparison between manufacturers' media was planned or contemplated, and if so, would you comment on it?

MR. MADDEN: I understand that the Environmental Protection Agency in their Cincinnati office is doing or planning to do a study. The definite study I am not certain of, but one of our people was contacted by the EPA and they did spend some time with us. And it seemed that gist of their whole thought was to carry on a side by side study, wherein they would have manufacturers participate in something like an NSF situation and determine whether on a hundred thousand square feet of manufacturer "A" RBC is equal or better than, or not as good as, a hundred thousand square feet of manufacturer "B" material. We have had a fair amount of experience in examining both your material and material from Autotrol and there is no question in our mind that there is a serious deficiency in the design and the availability of air to get inside the closed media, and it is something that has to be addressed. It is something the gist of the presentation here was to say look follower, we do not agree with supplemental air. If you get up in loading rates of 10, 12, 13 pounds soluble applied per thousand square feet of shaft, there is a trend at that point where you could look at supplemental air. In the case of RBC plant already discussed, they had a tremendous amount of odor problems in the initial start-up and one of the panaeceas suggested was to put in supplemental air. And they did put in supplemental air and it did not really help except it changed the air from inside the plant and pushed it ouside the plant. So, there was a lot of flack and I believe there are some people here who are involved in that facility, and they could even comment more than I could because you know we are a manufacturer and we look at the problem, we and when we can as Autotrol does, and as your firm does. we will go back and we will say look, this is what we think you should do if it is a problem that is outside the scope of what we could categorize as our responsiblity. If, and we all have had them, if you have media problem or you have a shaft problem I think it is certainly a credit to the manufacturers in this industry that we go back and sweep up our mess. And the important thing for you all to understand is that just because we have had some shaft problems and just because we may have different opinions about the way things are done, the RBC process is in fact a neat process. It is so good that I sincerely believe it is going to take over activated sludge in a very short time, as long as we can keep the equipment cost down and maybe keep the brass knuckles off between the manufacturers.

MISS KINNER: I was wondering if you had done any culturing or any microscopic observation on the films that you had there.

MR. MADDEN: No, we did not do that, Nancy. It would take people of your caliber to do that kind of study, and unfortunately when we started the study in all sincerity, it was sincerely a marketing study. What we were trying to find out was, was there a difference in the magic aurora of supplemental air, and quite honestly there is not any. We do not see it and we do quote project and design project based on non-supplemental air on a side by side basis with supplemental air, with air drive and we do believe the non-supplemental air system can be just as good as the latter system.

DR. SRINIVASARAGHAVEN: You kept saying that there is no difference in soluble BOD removal at Alexandria with and without supplemental air.

MR. MADDEN: Excuse me, no, I said there is a difference. That information was reported to us through you and our Mr. Friedman.

DR. SRINIVASARAGHAVEN: Right. The information that you got was not done on a continuous basis, it was probably taken from spot samples through your program established. The best way to compare the two would be to have a separate clarifier taking the supplemental air units through settling processes whereas the mechanical unit effluent went through the clarifier and then the total BOD removal itself. I think if it is done on a separate train basis you take the supplemental air effluent through the clarifier you take the mechanical unit effluent to the clairfier the effluent BOD. That process, I am certain, will be a difference in the total BOD coming out of the clarifier. Another point that you are talking about in terms of loading, Alexandria is designed as a very high loaded ystem, which you may be aware of. The design loading was ten to eleven gallons per day per square foot. During the pilot test the loading was the order of sixteen pounds of BOD_5 in the first stage and therefore the supplemental air effective in Alexandria because of the high loading conditions. For low loading conditions the situation may be different.

MR. MADDEN: I agree with you. The only point I would like to make or restate is that when we took an absolute comparison of supplemental air, no supplemental air, for our equipment we can look at it and say there is two absolutely similar or absolutely exact comparisons of wastewater and performance, and you can draw a conclusion from that. It is very, I agree with you, that it is bad to pick up Alexandria, pick up a piece here pick up a piece ther, mix it all in a bag and say well this is the result, and that is not what I am trying to accomplish here. What I did want to show is that when we took information that Autotrol published through the Purdue

Industrial Waste Conference, we wanted to show that there was no benefit to air drives, and again, I go back to what I said to Nancy before, that this was first and foremost a marketing look. Was there a benefit of going to air drive, was there a benefit for us to research and develop our own air drive and with the problems that we see with air drive such as rotational speed control problems followed with increased flow through a closed media system, we are not convinced of the worth of it. And that is what I am trying to tell this very profound group here this morning, that you have all seen really one side of the picture and what I am telling you is the other side of the picture, and it certainly bears looking at.

MR. DIAPER: I have two questions. Did you in any way measure the thickness of the film and do you have any restriction on that thickness. And secondly, did you try the effects of speed change on the growth on the media?

MR. MADDEN: First question, no, we did not measure it like Autotrol by weighing the shafts. We observed it, and on any given day, Tony, I could go and look at it, or one of the guys who was doing the study could go and look at it and could make an observation. At times we tried to call something yes, there is a little bit of a lighter growth associated with the air drive, and we can draw that conclusion at this study. But you would have to look very closely to pick up any discernible difference in the growth thickness or the growth characteristic and as we mentioned earlier, perhaps a microscope study where we did in fact look at the higher life forms to see what was actually happening there, that could be initiated, but with the aerobic growth which was dissolved oxygen not sinking through the second stage, if you notice there is an increase in dissolved oxygen from the first stage to the second stage, without supplemental aeration. And that kind of says what is happening in the RBC process associated with our open media. There was no quote sag in dissolved oxygen wastewater. We have never with the exception of one plant, experienced slides that you have seen here this morning concerning the Beggiotoa growth that I have seen. Now I have seen these types of situations and when I look at them, I say holy smokes, you know if that happened to our stuff I would be very concerned.

MR. DIAPER: Do you have any restriction on the film thickness?

MR. MADDEN: We have designed for a quarter inch over the entire media surface area. Now, we have not yet ever had that type of growth with the exception of maybe one location where we did have what was estimated to be three sixteenths using the same method that was mentioned by Mr. Sullivan, i.e., you take the surface area, take the weight of the machine and calculate out by process of assuming that the density of the growth is 62 pounds per cubic foot. You had another question, too.

MR. DIAPER: Did you try the effects of speed variation on plant performance?

MR. MADDEN: On one project we have two speed motors and we did not try at Cold Springs facility. Cold Springs facility does have a variable frequency drive system that we can change the speed and quite honestly you know, there are nine thousand things to study. What do you study next? And one of them would be speed and that is going to be an ongoing study for us; the speed versus supplemental air. The project will run through two experiments and we will have an accumulated data of the effects of temperatures. And there should be some interesting results which we will present at a later conference.

DR. SRINIVASARAGHAVEN: What are the loading rates employed and what is the cover effect on RBC?

MR. MADDEN: Well, soluble BOD, three to five pounds per thousand sqaure feet of surface area at first stage loading, lip to sometimes, now one facility is taking something seven. The other facility took 8.8 pounds per thousand square feet of surface area. Well we load Cold Springs up to fourteen pounds, both sides. There was no Beggiotoa none whatsoever. Another thing that has to be looked at, and perhaps some of the people like Mr. Sullivan who is standing up in the back patiently waiting for a question, just a minute, one of the things that perhaps should be looked at is what effect do covers have on Rotating Biological Contactor Process. And I will give you a case in point. Somebody visited our plant installation, and that is a twenty-four shaft job, it is in a building that has a relatively small cubic footage in terms of what you would normally expect for an RBC building but it is fairly tight. And in that facility, these two guys got busy. And one says I have to go out so he went ouside. And it suddenly dawned on these two guys that there was no oxygen in that building. And now there is a huge sign on the front door, "Do not enter unless the fans are running". Now one of the things we are going to look at, is what is happening on your cover; are the covers going to be oxygen limited, is there going to be sufficient oxygen under the cover to keep this process going? You are absolutely right. Humidity can be horrendous. I have seen some of our own plants, and some of Autotrol's facilities and some others, and humidity control is a problem. And that is one of the reasons why design consultants go to covers because they eliminate this problem with air transfer of heating, cooling etc. And condensation occurred inside the building.

MR. SULLIVAN: There is a number of parameters that you have on the board there. I think it is a very, very important problem. Your making statements then with regard to biofilm thickness, is critical. To define that, biofilm thickness at room temperatures at which the organisms' growth is critical. The soluble organic loading that you quoted for Autotrol's design conditions a three pound soluble. The actual data that has been gathered shows a zero water break point four pound soluble. Even that design is zero water dissolved oxygen limiting break point.

MR. MADDEN: Excuse me, let me stop you. I forget the question or the statement. We did not find that. We see a trend in that direction. But we did not find at four was zero. We found that we were still achieving not quite first order reaction beyound four pounds. But okay, that is a significant point. We do not agree with you there. And that is just a question of that, this is ours, this is yours.

MR. SULLIVAN: What are the other things; how long is the data that you gathered?

MR. MADDEN: Since June.

MR. SULLIVAN: And how many plants has it been based on?

MR. MADDEN: Well the pilot study done at Cold Springs is one plant. I mean that is just, one-two-three-bang, that is all.

MR. SULLIVAN: I think it is advisable that a research of the existing installations on RBC, both from a carbon removal standpoint, temperature standpoing, carbon and nitrogen standpoint, be evaluated. Some of the papers being presented here with regard to Ft. Knox, with regard to Alexandria, with regard to influent problems, not necessarily generated on the media, with regard to biofilm thickness, all play a role in the design of RBC systems. I think all of this information has to be correlated in order to come up with a proper design. The organisms have to be studied, biological changes, effects of DO, effects of slime thickness. And from a standpoint of pro-motion of the RBC indsutry I think this is critical. Yet when you gather data, it seems to me that we do have to take biofilm thicknesses. We have to know the types of biology present, we have to know the effects of temperature, we have to know the desired oxygen concentration in the reactor, we have to know the biofilm thickness.

MR. MADDEN: I agree with you. I think all that is very important. I think though that most of that is important for you and for us and for other companies that are promoting this equipment to basically consulting engineers who have the same problems you and I have. We have to meet a payroll every week, we have to get the drawings out, we have to contend with the EPA and for the group that is here from the academia, they want to know what kind of a bug is living there. All they want to know is that when Autotrol, or Clow or Hormell, or other brands say that this is what it will do, that in fact they will feel comfortable with that. And I think you are right. It is incumbent upon us as manufacturers to do the work and I did not intend that this should be the end-all be-all for RBC Supplemental Air. But one of the things that led us to the study was in looking around at our operating plants. We do not see this kind of problem. So then we say well maybe something is not hitting us on the head and that was the reason we started with this study. And as I said earlier, since your firm was enamored with the air drive and possibly because a lot of consultants do like adding a little bit of air in the wastewater treatment system, hey that is the best end of the world. But if that is the end of the world perhaps since Clow sells aerators too, make the tanks bigger, throw out the RBC and put aerators in. You know it is the other side says we cannot see 200 CFM of air, the manufactured at the inefficiencies in a blower system clumped in the bottom of a tank and quite honestly, you made us look at it and we are still looking at it.

DR. CHOU: I am in response to you mentioning the soluble BOD removal at South Shore. South Shore has about one third to a half of the BOD contribution from the industry, a single industry which is animal group. We found these soluble BOD at South Shore less than the average that we have found for domestic waste. And also we found that the soluble BOD removal could reach even higher than ten pounds depending on what type of wastewater you are talking about. So there is one factor I think you have missed, which is the waste characteristic in response to waste treatability.

MR. MADDEN: You are right. We studied one facility, you also studied one facility. In that facility, you do not state any place in the publication at Purdue, what actually is happening.

DR. CHOU: No, we stayed at South Shore, exactly, and we also mentioned that load from the industry is one third to a half. Because we have to get the paper accepted by putting in the industry.

MR. MADDEN: Well, I understand. But, one of the things that you are missing is, and we recognize it, I am sure you fellows recognize, is that different wastes have different biological oxidation rates, and that industrial wastes can move the growth farther down the flow-path, or move it up closer to the front, hence a better reduction or a better reaction occurring on the media. We recognize that and again I will state, this was one study, one look at one location. We are continuing the study but it is worthwhile in our opinion to say that we do not, we have not had this experience problem that made such a strong point with me for this morning. And I do expect that we do have to respond for it.

DR. CHOU: Yes, we think it is very important to address to the audience because if we only mention the one side of the fact we could be misleading. And in regard to the oxygen transfer, you mentioned that the structure, the internal structure of Autotrol's product could be inferior to yours in regarding to oxygen transfer. I would like to ask what is the percentage of oxygen transfer to the liquid phase, versus the oxygen uptake in the air phase, because this is a three phase, involving the solid biomass and the liquid phase and the air phase. So have you done any work as regard to the oxygen transfer in each phase?

MR. MADDEN: No we have not, and I do not know, but let us go to the other side of that. We have examined the bundles and what is growing there. And we have pulled yours and we have seen what is growing or not growing on yours. So all I can say is that there appears to be a difference and that is why we are looking at this difference and we are going to continue to look at it.

DR. CHOU: I would like to address some data rather than the appearance because your biomass looked to me like Beggiotoa - all white. It is not black but it is all white.

MR. MADDEN: The basic media itself was white, that is to say that the polyethylene itself prior to the time we started using carbon black for U.V. addition, it was absolutely white, bone white. And the last slide that you saw that was nitrogenous bacteria and it was a white media, so I cannot make it what it is not you know, there it is.

DR. CHOU: Now it is my turn to present a data in terms of part of oxygen per horsepower per hour in comparison with the surface area. I am trying to make the point that the dictating transfer is in the air phase rather than in the liquid phase. We have done some work on the liquid phase. In response to your question, your mentioning of the ethereal structure of Autotrol's product, I do not think that is a legitimate attack because we feel that most of the oxygen is transferred through the air. And there is no limiting as far as oxygen diffusion in the air. So it does not make any difference what type, you can use a flat disc or dimples. One more contribution! That is one thing we have done. We also have learned from Japan, the enclosure oxygen-limiting condition. They have done work on O.R.P.. This is window area, to define what is the requirement on window area to avoid the oxygen-limiting condition. So we have data for that.

Session 6. INDUSTRIAL WASTEWATER TREATMENT
Presiding: J. B. Walasek
Wastewater Research Division
U.S. Environmental Protection Agency

DR. PAGORIA: A couple of questions please. When you say 10% seawater I just want to be sure I understand. Do you mean a hundred parts per thousand salinity or do you mean approximately three times the strength of seawater? When you say 10%.

MR. LANG: I mean 10% of the salinity concentration normally found in seawater.

DR. PAGORIA: So you are talking about 3.5 parts per thousand. That is good. the temperature decline was caused by what; evaporation?

MR. LANG: Yes, temperature change from 28 to 30 degrees at sea level.

DR. PAGORIA: And then one last thing. The source of nitrogen, I am sorry I missed.

MR. LANG: One four.

MR. CHRISTIAN: I would like to know what pH measurements here 10% salinity corresponded to.

MR. LANG: The 10% we have pH of 6.5 to 7.8.

MISS KINNER: What was your hydraulic loading on that?

MR. LANG: The hydraulic loading we did it in meters of 100 square meters per day. That was 6,100.

DR. POON: I could not see the data very well. I wonder if you have been able to show if you have increased your ammonia nitrogen loading. Would that compensate some of the inhibitory fact of this chloride? I seem to find in the removal of organic relationship that seem to exist. For example I increased the BOD loading to RBC, I could reduce the inhibitory fact of chloride from the removal. Now does that apply to ammonia?

MR. LANG: The effect of high chloride concentration on organic and nitrogen removals in RBC should be similar to the case in the activiated sludge process. However, we did examine this effect in our study.

DR. JENKINS: What was the type of ammonia in these systems to growth nitrifiers?

DR. REYNOLDS: There was a pretty good nitrification reduction and that may indicate some problems with the Ks and u_{max}.

DR. JENKINS: I was going to suggest if nitrification were taking place you would expect it perhaps largely in the end of the disc system which might account for your inability to develop satisfactory growth constance based on COD or BOD removal.

DR. REYNOLDS: I think you are right, Dave. And I appreciate that comment. One thing I ought to point out though Dave, for your information, cheese wastes often are nitrogen limited and we had a concern about that in this particular case. We did do some C/N/P ratios and it was indeed, carbon limited. Your comment is well taken and I appreciate it.

MR. ATHAVALEY: When the time engineer look at your equation, you are referring to constants which pertain to the particular substrate for example COD as discussed here. Now, if I want to know what would be BOD and nitrogen effect on these constants and if that is the case, are these constants interrelated or could any attempt be done in that direction to achieve those number of kinetic constants.

DR. REYNOLDS: As long as you were talking about the carbonaceous removal it is a fairly easy transition to go from BOD to COD or to TOC. That is the coefficients and the constants are easy to come up with. They are, if you will, just a fudge factor away. When you try to go from carbonaceous to nitrogenous removal, in other words nitrification, then a whole set of different parameters would have to be developed.

MR. WONG: Have you done any toxicity test for the RBC effluent to pond?

DR. BRACEWELL: No, we did not observe any toxicity in the pond systems so we did not actually have the time and money for us to do a toxicity test by ourselves.

MR. BECKMAN: Where did you collect the samples?

DR. BRACEWELL: We measure total TOC on the influence and effluent waste stream.

MR. BECKMAN: Do you have any idea what the feedbacks are going to be on this gas evaporation system? Are they going to be better in four or five years?

DR. BRACEWELL: They currently incinerate now. But compared to what the current costs are. It will pay back within four to five years.

MR. BECKMAN: What are their thoughs for that now. Do you think it will be less?

DR. BRACEWELL: It is possible, nothing sure. Although the cost is going up more it can be paid back within four to five years. Originally we were so upset, they were going to have the natural gas supply shut off.

MR. ANDERSON: Can you comment on your method of sampling?

MR. WATTS: Well we did take what were the samples available. We were able to get one what was a sample because it rained on. The total solids most of which are total dissolved solids from the process. The soluble contribution from that is not that. There may be a certain amount on the process path of separable organics that will go down the sewer and there is a lot of separator as an interval part of equalization basin. But from the present knowledge that we have, those things would be on the process pattern up to decide and if they were, they would go down the sewer and their part of the material balance. In other words say there were more wash down and things like that in material balance in the original dry water data.

Session 7. CONCEPTS AND MODELS
Presiding: R. D. Miller
U. S. Army Environmental Hygiene Agency

MR. HYNEK: I noticed your logic in reducing the amount of surface area in successive stages and I believed you said correctly a fifty percent reduction. And then you have said in the trail which provided successive less oxygen. However in your case you have cut time, in their case they have not. You have indicated no recognition of the changing degradability factor in a complex waste like municipal waste. The easy stuff goes fast, the last five parts go slow. Could you comment on that please?

MR. STEINER: The reference to the Minneapolis application was for the exclusive purpose of showing the disc as a mass transfer device. That is to say that as more treatment is achieved there is less aeration required in the successive stages and therefore, the conclusion was made that with respect only to aeration that the succeeding stages could be made smaller. The intention was not to compare the residence time with any other factors. With respect to your comment on the decreasing Q value or the lessening of treatability in successive stages, I certainly agree that the top BOD comes out a lot easier than that which was remaining. If you were to plot however the data generated at a place like Pewaukee, you are not getting down to two, three, four and five parts BOD you are up a little bit higher than that. You will find that the existence of refractories simply does not make itself known when you go through plots of data. I do hope it is heredity of the data does not seem to get to the program correctly.

MR. HYNEK: You have used the applied ammonia per thousand square feet with indication of over tweny-five pounds. Is this correct?

MR. STEINER: That is correct.

MR. HYNEK: In your paper, the units are somewhat off and I would request or make this question; do you have data to back up your position or as to your logic?

MR. STEINER: Yes and I will refer to Dr. Borchardt's EPA paper for the data that will substantiate the curve. Also the data for these from Chicago study on nitrification as well as others.

MISS BERGS: You said the RBC design is based wholly on organic loading. Is there any minimum of retention time that you would have to include hydraulic loading?

MR. STEINER: Our work with detention time was well over one minute or less than a minute I cannot say. But the detention time was down to one minute. I have found no significance of detention time with respect to removal down to one minute detention time.

DR. O'SHAUGHNESSY: I agree with your statement that the RBC treatment plant design should be based upon organic loading. However, for nitrification, our

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early work indicated that decay of ammonia nitrogen as defined is ineffective. But when you change your surface to volume ratio on your reactor it will have an impact on nitrification kinetics.

DR. WU: I would like to learn some basic knowledge from your excellent work. What you have mentioned in your presentation is the attachment of the cells on the surface of biodisc is due to the formation of polysaccharides. If this is true, how do you make a suggestion to the process design engineer in regard to the optimization of cell attachment in the fixed growth RBC system?

DR. LAMOTTA: We do not know yet. We are going to perform more research work in this area so that the optimization can be achieved. However, the objective of this paper is to study the substances which can be used to precoat the disc surfaces. As a result of the application of surface coating agent, we can get an immediate cell attachment and the acclimation time required for biomass to develop on the disc surface will be significantly reduced.

DR. WU: As far as the effect of magnesium limitation on cell properties is concerned, the other studies indicated that the production of cell polysaccharides increased as the degree of magnesium restriction also increased. Can you comment on your experimental results that show the cell attachment on the disc surface reduces as a result of the reduction of magnesium content in the wastewater?

DR. LAMOTTA: You said that the addition of magnesium into the wastewater is not going to improve the production of cell polysaccharides. It may be true but however, the addition of magnesium in the present study is for the enhancement of cell bridging and/or attachment to the disc surface.

DR. UNZ: The addition of magnesium and calcium to the wastewater is a good idea in terms of the improvement of cell attachment. I think calcium is much stronger than magnesium in this case.

DR. MUELLER: You have got a real nice definition and I agree with it a hundred percent. I guess it is a philosophical question of when you take existing knowledge perceptive and apply effectively to get what you can out of it and when you continue to do research. I think applying the first principle we have today, making some assumptions about the reaction coefficients, and I am sure I will show you again this afternoon the additional knowledge as to what some of the important factors are and where those research efforts should be directed.

DR. GRADY: I cannot say that any better, Jim.

DR. POON: I have two quick questions. One is concerned about the addition of lime to RBC sludge for the improvement of solids settlability. Does the chemical addition impair the sludge digestability? Secondly, do you have any problem in the winter, any calcium deposits because of so much lime in it?

MR. NOSS: A friend of mine is now doing a study on the digestion study of the sludge produced. He is also going to be incorporating into the paper the amount of sludge produced, the characteristics and dewaterability of RBC sludge.

That paper will probably be coming out about the end of this year. But I have not attempted to put any of that data into this particular presentation. However, it does not seem to be any problem with handling the sludge and whenever our sludge production is not all that great compared to what we would normally see, in other words, with the increase in the amount of secondary sludge produced when it is mixed back with the primary sludge we did not really have any problem in handling it. To answer your second question, we have no problem with calcium deposits. We did not see anything falling out into the RBC. To be truthful about it, the use of lime to form the calcium carbonate which put down the phosphorus and the appetite in the primary clarifier. We have no evidence of any residual built up in the RBC system. We will further investigate this to see any build up before the end of the project.

MR. SMITH: What temperature range was used for your experiments?

MR. NOSS: We ran this particular experiment inside one of the buildings at Fort Detrick. Therefore, at lowest temperature was fifteen degrees because the building was heated in the winter. The temperature range was between fifteen and twenty-five degrees.

MR. MEANS: In the first part of your presentation when you talked about feeding the flow into the four trains; do you know the application of alkalinity was prior to the first train and after the last train in the pH change from 9.5 down to about 7?

MR. NOSS: We did the standard calculation on alkalinity destruction with respect to ammonia removal and we measured alkalinity ranging generally from 200 to 250 mg/l. We were seeing essentially the magic number of seven miligrams per liter of alkalinity destroyed per miligram of ammonia removed.

MR. BACHTEAL: From your experience, those which you cannot write about, but at least you can talk about in the abstract, would you prefer to see staging or the loading rate such that one module would not handle the load? Would you prefer to see two or three modules or four modules in one stage or alternatively to split the flow in two, three, or four streams such as you fed 25% of the flow to module A which would be a first stage module and 25% to module B which would also be a first stage module?

DR. FRIEDMAN: I have mentioned that question before. My initial reaction to that question would be I hate to put all my eggs in one basket so I would be in favor of splitting the streams. If something goes wrong with one of them, we do have and will have in the future drive problem, I would rather not have all my eggs sitting and waiting on in stage 1 and they should be distributed to stages 2, 3 and 4 behind it.

MR. SUTTON: Can you explain a little further your current data that you had on influent versus effluent in long term BOD and you related that to an existance in the effluent BOD concentration much greater than the influent. And I believe you were relating it to hydrolysis occurring in the reactor and therefore getting a BOD would not show up in the sample of the influent.

DR. FRIEDMAN: This is a particular wastewater that is incredibly high in protein but as such as the total protein which I call the readily biodegradable, I was very careful to call that when I started out. That disapears very quickly from the RBC. No further treatment or reduction of either COD or BOD_5 with the rest of the six stages beyond second stage was found. Subsequently, because of this we had very strong evidence but not complete. We ran a complete protein test and the result of our lab tests showed that at least 80% of soluble COD was protein. It is obvious from this study that we are not breaking down all protein despite every opportunity to do so in the RBC system. And yet it did break down under a long-term BOD test. That is the only part I was trying to make. Eventually it is going to break down and it is going to be in the receiving stream.

MR. SUTTON: Why did not it show up on your effluent sample as well? The influent BOD sample did not behave the same as your effluent. I do not understand.

DR. FRIEDMAN: This might be what you would call a refractory component of that. Refractory is not a good word in terms of BOD but a more difficult portion.

MR. SUTTON: You feel that it is because of the shorter retention time of the RBC plant that dictates against it in this case?

DR. FRIEDMAN: Well, it was not talking about the same kind of the wastewater and running into a suspended culture system and giving a very long mean cell residence time. My personal opinion is I have no proof of it, that after the protein bonded on the cell surfaces, it takes awhile to break them down. When we have a large amount of sludge with a very long detention time it has an opportunity to do that. When you take the same concentration of the wastewater and one of these systems you do find the breakdown of this protein.

MR. IANNONE: You briefly went over the carbonaceous part of your talk. I was wondering if you feel that the carbonaceous portion of your model is verified at this time, and are you designing systems or would you design systems, in accordance with your model prediction? Before you answer that question we have taken some of your data and compared it to existing design curves and you seem to continually fall below the predicted curves.

DR. MUELLER: Yes, I would use this to design and this is the only way we do design RBC systems. However, our analysis is still a specific plant and still a specific waste. On the initial carbon work was done on industrial wastewater two different types of paper mills. We also did it for domestic wastewater and we got a different K rate. The K rate is a function of the wastewater we are dealing with. The nitrogen work already anticipates that the K rate or growth coefficient, the nitrifier, should be constant. The growth coefficient with the carbon will not be constant. That is a function of the wastewater. As Dr. Friedman mentioned we do not have a system here which we can just take a design without putting some limits on that design.

Our results in the model simulation vertified the results for that giving wastewater. I certainly wish anyone in this room to feel they can take my result and use it for designs in their system. If I do not have data on your system and a K rate from your system, I cannot use my model unless somebody uses somebody else's K rate. How does that simulate your rate? It beats me.

MR. IANNONE: Do you think the K rate for domestic wastewater is going to vary that greatly?

DR. MUELLER: To a large extent, yes.

DR. LAMOTTA: Does the kinetic behavior of the biological solids differ between the fixed growth and the suspended growth system?

DR. MUELLER: Yes, we have noticed some differences in kinetic behavior between the fixed-film RBC system and the suspended-growth activated sludge system. According to our experience, the kinetic coefficients resulted from carbon work differed only slightly but a remarkable difference was found in nitrogen work.

Session 8. UPGRADING PRIMARY AND SECONDARY WASTE TREATMENT
SYSTEMS WITH RBC
Presiding: B. J. Garg
Department of Environmental Resources
Commonwealth of Pennsylvania

MR. SULLIVAN: Just some basic questions with regard to clarifier efficiency. In the course of the evaluation was a long-tube settling test conducted? The clarification rates for secondary solid separation that you suggest in the paper on the order of 500 to 600 gallons per day per square foot; recent publications have shown around 1500 to 2000 gallons per day per square foot to reach effluent 30 solids. Could that be a function of the turn-around section, or were there two settling tube tests conducted to determine actual solids drop velocity?

DR. MUELLER: In that section which as you know was taken into account in calculating overflow, the actual overflow rate was about 500 even 450 at this point, which is varying slightly. They did conduct a tube settling test, to generate that curve but the data that we applied to this was the data from the five main field tests, that would be out of solid data from the field testing itself.

MR. SULLIVAN: I guess the question is, is there significant difference from the actual field test on solid settling velocity to long-tube settling test, and could that be a function of clarifier turn-around sectional modifications of the physical configuration of the system?

DR. MUELLER: I do not know if there was any difference between the two and what they found. I presume that but I cannot agree with. Again I think you put the very long tub properly and we knocked out 25 percent of the remainder. Why the solids did not settle as well as what you found in the field I do not know. There is a piece of data which represents the result. Theoretically, except for that the clarifier function is almost based on solid settling. I really cannot answer that what is the apparent difference.

MR. SULLIVAN: One other question in design basis. Your final design selection is based on a 130 soluble BOD. What was the input from the industrial applications on that?

DR. MUELLER: That is a good point, I have not got the foggiest. You have an idea on that?

MR. SULLIVAN: Well, basically what we are doing is designed into a peak monthly and because of the industrial application out there you have a high organic surge that is about seventy percent during the peak month over the average month and hydraulically about fifty percent during a peak month over the average month. So your design basis is about 2.3 times the average monthly condition and the economics reflected peak month design that is 2.3 times the surface as a result forces you out of the existing primary tanks.

DR. MUELLER: I agree with you yes, no doubt about it.

MR. SULLIVAN: And it has a significant impact on economics, factor of two.

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DR. MUELLER: No doubt about it at all, I agree with you a hundred percent on that.

MR. DIAPER: Jim I enjoyed your paper on Edgewater. Was this the installation where we did the micro-strainer tests?

DR. MUELLER: Not in conjunction with our study.

MR. DIAPER: I remember the consultant Dick Tolbert rented a small unit from us and took it up to Edgewater and ran some tests downstream of the RBC. The point I wanted to make was that if you were looking for a solution to the removal of solids the micro-strainer would be a very effective on because the RBC produces a stringy type solids that is easily filterable. In fact we have got flow ratings of twenty GPM per square foot, and drop solids from a hundred to the low ten on a single pass. The installation of a micro-strainer there for 3.6 MGD would need about two ten foot diameter by ten foot long machines, and the interval costs would be about a hundred thousand dollars.

DR. MUELLER: That is a definite possibility, instead of going with say a sand filter on the end to get the effluent quality you do a micro-strainer. You might be in better shape.

DR. SRINIVASARAGHAVAN: The ammonia and BOD loadings throughout the study have been pretty low compared to many of the other studies and I understand why. But that kind of a loading where there is absolutely no limitation oxygen transfer especially in a small scale unit, should there be any consideration for scale-up at all? Why does the scale-up factor? And the second question is you said the soluble BOD may not be the best parameter to monitor the RBC performance. So I would like for you to comment on this BOD consideration for RBC design also.

DR. POON: Well, first on scale-up, I really have no...I really am not sure what should be the scale factore. Therefore I said if the manufacturers suggest a certain range I would feel confident that you can go down to the lower range off their suggested scale factor. I do not mean to suggest this would be the scale factor recommended to everybody. And the second question now is concerning my question on the soluble BOD parameter. Of course the reason for people using the soluble BOD is that they consider the RBC only removed soluble BOD and the particular one would probably go through intact because the removal would be more effected by the clarifier in the RBC system. Also a look at my data, I thought I might have a transparency at hand to show if you take grab sample when random sample daily, a lot of times, a lot of particular BOD in the trickling filter effluents coming out and go through the RBC system, but they do not go through the system intact. And sometimes the particulates BOD incorporate into the biofilm, but a lot of times there is a lot of sloughing out of the biofilm contribute to the particular BOD in effluence before it reaches the clarifier. So I have data to show many times negative value, in other words no sloughing at all, and actually build up biofilm look at that particular sample. And so I have negative value, zero value and positive value for sloughing. But what I am trying to say is particular BOD does not go through the RBC system any time. That is the internal changes within the system. So I think that soluble BOD may not be the best parameter to monitor the performance of the RBC.

DR. SRINIVASARAGHAVAN: The plant performance has a correlation with the BOD; would it be any better relationship between the performance and the loading?

DR. POON: I do have the curves using total BOD. I will not say it is better but I could use that data for the design also.

MR. STRATTA: I notice you presented some data on mass of biofilm per square centimeter. Will you describe how you got that data, scrapings or what have you, and where it came from the discs?

DR. POON: The method is simple. Just scraping off the amount of biofilm from the known surface area and dry it and we weigh it.

MR. STRATTA: Was it representative of the disc?

DR. POON: I would not say it is representative of, that is why I say we do the best. I notice that the biofilm is not even within the same stage and as I have listened to many speakers that in the modelling of the RBC they assumed uniform thickness throughout the RBC and that never happened. If you go through the RBC facility you do not see any uniform thickness at all.

MR. STRATTA: I listened to your paper with great interest. You did not mention much about suspended solids. Were you trying to reach a standard on suspended solids?

DR. POON: Our suspended solids concentration is quite low, most of the time below 100; I would say between 50 and 100.

MR. STRATTA: Were you trying to reach a standard on suspended solids?

DR. POON: I was saying that the average suspended solids coming out of RBC clairfier was 15 mg/l.

MR. STRATTA: Fifteen?

DR. POON: Yes, so I allow under the worst part of a siuation another 15 mg/l BOD contributed to the final effluent.

MR. DIAPER: The other question I had was I saw your picture of Philadelphia and I heard that there was a problem with Philadelphia. Was there a failure of several shafts?

MR. McCANN: There was some problems with the shafts at the installation, I believe one or two of them had problems.

MR. DIAPER: Could you expand a little on the nature of the problem?

MR. McCANN: This question has been raised earlier in the conference and I think we addressed it at that time. There was a problem with some of the shafts. We think we resolved the problem and our testing shows that the shafts' current design is adequate and there is no problem with it. The City of Philadelphia's solution, they have resolved the problem in their own mind and they have no problem about the use of the equipment.

MISS KINNER: Do you have any problem that has an RBC system before and after they have activated sludged?

MR. McCANN: I believe we have one or two installations. And when you say roughing, what usually happens is that the nitrification upgrades and the roughing becomes a symmetric because what usually is required is secondary treatment and the activated sludge is converted over into a nitrification treatment system.

MISS KINNER: If you got your need in nitrification, would there be any problems in our loading the RBC in activated sludge basin?

MR. McCANN: No, the system would be designed, the engineers' requirements would usually be that the effluent from the RBC system be such that it would provide sufficient waste to maintain the activated sludge system. In the RBC system design we would attempt to maintain the loading so you did not have an overload condition. So it really depends upon wastewater condition.

MISS KINNER: Are those two things compatible?

MR. McCANN: Yes.

MR. WARD: What I am wondering is whether you have had any experience following your RBC process with a tube settling module. This is what we are doing currently and we have come up with a situation where the RBC has the selective organism which adheres tenaciously to plastic. This follows through to the tube settling modules and we get a very heavy growth which eventually begins to ferment and produce gases and cause the modules to rise to the surface. I was just wondering if you have had experience with this type of a problem in other installations?

MR. McCANN: Again, the tube settlers for whatever reasons are seldom utilized in our installations or I would presume in most biological treatment systems. We have done some experimental work with it. Usually the tube settler is utilized for a fine tuning of the effluent. Usually it comes from a secondary clarifier and you normally will not have the heavy growth build up. Once you start to develop the growth some means of removing it, usually just dropping the liquid level, causes the growth to slough off. Our experience has been that the growth is not that tenacious as far as its clinging to the settlers.

MR. WARD: We tried a combination of chlorine application immediately before the tube settlers in order to reduce the growth in addition to periodical flushing and it is such a strong growth at a speed that we can drop our tanks while there is not enough to provide a shearing force to really do any good to begin with it. And I was just wondering if you had any other...

MR. McCANN: To my knowledge, again, we have only utilized the tube settlers following a clarification procedure so that the loadings you are talking about, the concentrations onto the tube settler are on what range? What range are you using?

MR. WARD: Entering, or a secondary clarifiers, we have a fifty parts per million but most of that is settleable, so that leaving the tube settler modules we have about ten parts per million of BOD or less and comparable suspended solids in a ten to fifteen mg/l range. But the growth that we get would amaze you at how thick and strong it becomes. Every piece of equipment following the RBC process. It is not a problem on the walls and the flight guard and so forth, and our clarifiers we do have a problem with the tube settlers and we were just wondering whether there really has been anything.

MR. McCANN: Not to my knowledge.

MR. ODEGAARD: I have two statements regarding the separation of the sludge. In Norway there are several plants that use microstrainer as the separation of the biodisc sludge. Is not the same type that you use here in the United States? And about the sedimentation or tube settler used for military installations sludge in Norway, and there have been considerable problems on disc sludge with Nomella. This problem has been solved by using a vibrator that vibrates very gently the Nomella.

MR. WALL: I noticed you have a plant in Ontario and one in Pennsylvania and one in Peoria, Illinois that had the fiber glass covers in sort of a winter climate. Do you have problems with the doors condensing and freezing shut in these cooler climates on the fiberglass cover?

MR. McCANN: We have not experienced any problems with doors freezing in cold climates at all. The interior of the ambient temperature underneath the cover usually assumes the temperature within one or two degrees of the water temperature so that there is usually no problems with a freeze-up in the cover.

MR. DIAPER: I am interested in the solids removal from the RBC. As I pointed out earlier the microstrainer has been used for that purpose. What was the system you used at Cadillac, Michigan?

MR. McCANN: Cadillac, Michigan uses a mixed media filtration system.

MR. DIAPER: Does that go onto the RBC without an intermediate clarifier?

MR. McCANN: The filtering consists of activated sludge followed by secondary clarifiers and then during the summer operation for nitrification RBC and then directly onto the sand filters.

MR. DIAPER: What sort of removal efficiency are they getting?

MR. McCANN: As far as the solids?

MR. DIAPER: The suspended solids.

MR. McCANN: The suspended solids come into the system usually vary only within one or two parts of the incoming suspended solids so they very seldom exceed ten milligrams per liter of suspended solids coming into the RBC and onto the sand filters. So the loading extremely low but the solids coming off the sand filters are under five milligrams per liter.

MR. DIAPER: Many years ago when Autotrol was first getting into this business, I did some microstrainer tests with Pewaukee and we had 100 mg/l coming into the microstrainer and about ten going out without any intermediate clarifier. I wondered if you have given any consideration to the combination of an RBC and a microstrainer?

MR. McCANN: We are not in the microstrainer business, so we are not usually really concerned with what is following our system. The engineer's choice is to what he will use. Any opposition using microstrainers is not ours.

MR. ANTONIE: The question that was asked before with regard to discs before activated sludge plant. We did not report on it because there was not enough data available, but there is a plant in Crawfordsville, Indiana with discs prior to activated sludge and there is a nitrification plant and very specific with regard to the soluble BOD desired to the nitrification system. In the capability of bypassing a number of shafts and actually putting on surface area against the load to reach an effluent concentration designed hydraulically. For instance, if you start up the plant half design you have the capability to use half the number of shafts or a fourth the design with three-quarters. So generally if you are looking for a design soluble concentration with the activated sludge system on suspended growth, you should hydraulically design flexibility into it so you can control the concentration.

DR. POON: When you use RBC in conjunction with your trickling filter, should it be put in front of the unit? In front of the trickling filter or after or in parallel? And if you do have an opinion one way or the other, what is the reason?

MR. McCANN: The utilization of the RBC is in a trickling filter application, usually the most confining of the two activated sludge trickling filters. To my knowledge, we are not, they are seldom if ever utilized in front of a trickling filter. You would usually not utilize a trickling filter or a final tuning of the effluent criteria. We would more likely to be utilized following the trickling filter for extremely fine effluent, usually either for nitrification or again for extremely low effluent criteria BOD. As far as the parallel series, we have not...again it would depend upon the hydraulic level. I have seen requests for both and seen them operated as both. And it really just depends upon the client, we have really no preference either way.

DR. SRINIVASARAGHAVAN: Could you translate some of the data plotting that you show for organic and ammonia loading rates in pound?

DR. MILLER: We did some of it at one time and the data is available. You could do that but, no I cannot right now.

DR. POON: I did some calculations. I am really asking two questions. Use of RBC we do not anticipate any filter fly but in my case I put the RBC right next to the trickling filter and the trickling filter has the filter fly. As a result, I have a lot of filter fly in the RBC system. Did you experience that?

DR. MILLER: No experience to problems with filter fly in the trickling filter. We have problems with sludge worms. We went over to get what was identified again, visible observations where there were sludge worms and actually this problem even came into the RBC process itself. At one point it looked like red meat balls in the RBC stages. I have not experienced on a full scale or not, but again it was a very short time in the heat of August with us, for about one month and it did not seem to have any actual effect on the process natural conditions.

DR. POON: What is your opinion of lime chemical addition to control pH and of increasing nitrification rate in the recarbonation system. In very cold regions would there be any problem with scaling at that position with calcium?

DR. MILLER: I really do not think so. Mr. Stratta of Penn State University, he is continuing to work on chemical addition. Currently he is working on a doctorate at Penn State and the specific thing he is working on is chemical addition in the RBC process.

DR. POON: I saw the slide in Dr. Friedman's presentation yesterday. I think in the recirculation area they do show a lot of calcium deposit on the first and second stage of the RBC. So I just thought that in a very cold region it could be a problem with lime addition.

Session 9- DESIGN AND OPERATION
Presiding: M. F. Saunders
Department of Civil Engineering
Georgia Institute of Technology

MR. OPATKEN: The value of K_{L20} is related to the viscosity and therefore that is saying that if we get an increase in viscosity we are getting an increase in mass transfer. I find hard to correlate those two.

MR. SEVERIN: As it turns out there are several ways you can plot this data, and I have been told that I should use a Schmidt number and you can manipulate a Schmidt number which is a Ki. Rather than getting into the nittygritty chemical engineering of this thing, I would rather point out the major lesson I think is that not only do we have to look at the different media designs but I think geometry is playing a very important role in the oxygen trnasfer.

DR. REYNOLD: Would you please get into a little more detail on how you came up with emperical dimensionless number that you were plotting that is the N_{20}. Why did you include the variable that you did as opposed to some other set of variables that could have been picked. This sort of relates back to Ed's question about fluid viscosity. How do you get your things together and you made it fit and I wish you would go into just a little more detail.

MR. SEVERIN: We found the N_{20} value as given by Dr. Wu in a recent paper of his in Water and Sewage Works about one year ago, where you see that dimensionless parameters to ascrible mechanical mixing and how he finds the oxygen transfer constant. So after finding it we were looking for a number that would be the dimensionless, and involved some of the parameters we were looking for and it happened that we saw Dr. Wu's paper and used that dimensional scoop.

MR. STUDEE: One concern I have had about a potential problem with the air drive is that it might be foaming problem in the RBC contactor unit. Evidently you did not have that problem here from what you have said, but I just wonder if you considered that and if anybody else in the audience who has had experience with that.

MR. HYNEK: Yes, we did. My concern with the RBCs back before we actually designed this plant, had to do with foaming was one of them, grease formation was another. That was part of the reasoning for aearting the primary clarifier. We thought by putting air in the primary clarifier you could remove partially the grease that might be in the wastewater and get it to come out in the skimmers. I was concenred a little bit about foaming particularly because the plant is underloaded. And wastewater has a tendency to foam more with addition of the air.

MR. JAFFER: I was curious if you have had any observations on rotational speed as far as eveness and how it might be corrected if it is uneven.

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MR. HYNEK: We have been able to adjust but it is not easy to adjust the balance because once you change one of these butterfly values affects all the other butterfly values' settlings. So it is a little bit of a frustrating operation to try to achieve any one particular RPM because it is a matter of continuously going along and adjusting the butterfly valves. At the present time through we have been able to, you have the second set of RBC, the rotational speed at the present time, I am just reading this to you, on the first stage is at 0.95, on the second stage it is 0.8 and on the last stage it is 0.62. On the first stage now which is receiving all the load, the rotational speed is 1.7 on the first stage, 1.2 on the second stage and 1.0 on the third stage. So we have been able to achieve that with a little bit tampering around actually with these butterfly valves. Once they are set though the thing remains exactly that way. We have not experienced an unveness, so that the things go, you know, flop-flop-flop; have not seen that at all, they run perfectly smoothly.

MR. HANKES: I am a little bit confused as to the data and the comparsion of horsepower drawn on the air drive versus perhaps mechanical drive. You mentioned units being taken out of service and operating at a slower speed. I wonder if these considerations are taken into accounted when making a comparison between mechanical and air drive?

MR.BERNER: The way we got the air drive on the particular plant, the plant was originally designed for mechanical drive. An the set-up was 7½ horsepower drive motor with pretty much of a standard Autotrol design. Autotrol came along and said we will supply the air drive for this plant at no increase in cost to you and furthermore we will guarantee the thing, so that if for reason our air drive does not work we will take the air drive off and put the mechanical drive units on at no additional expense. From our point of view that looked like a pretty good deal because we had an irritating problem trying to get the molds for the mechanical drive above the flood elevation. We had long chains, and it was just kind of a micky-mouse arrangement. So the air drive was a realy advantage to us, it did not cost us anything. On the 40 horsepower motor that drives the six RBCs, we have actually throttled that down to a place now where it is pulling 33 amps or 25 horsepower. The full load ampage would be 52 amps. That is 4.2 horsepower per shaft. And that does not count--- we have got two air headers going from that single blower to aerating the primary clarifier, so we do not exactly how much air that is. However, we think we can get this down to 2-3 horsepower per shaft.

MR. HANKES: Part of my confusion comes from the fact that yesterday we were discussing, I believe that the required horsepower on mechanical drive on the order of 3-4 horsepower operating at 1.6 RPM. We also mentioned that the horsepower versus feed was perhaps an exponential realtionship, and with these units operating at power would be that you are comparing against.

MR. BERNER: Well I do not know what the actual mechnaical horsepower would be. I think my main single concern with the air drive unit would be the restarting of a unit which had somehow sat for a long time maybe because of power failure, and I think that a lot of the horsepower that is in the mechanical drive units is put in there just, you know, to be able to restart the thing if it happens to be down for a period of time. But the disadvantage of that course is you to pay for that horsepower.

MR. HANKES: One final question. Do you have provided any supplemental means for starting the units other than air, if they are down for an extended period of time?

MR. BERNER: I think you were down for a long period of time we would have to reset the system and restart the system again. We do not have any other supplemental means.

MR. HOEFLE: Did I hear you characterize a shaft length as twenty feet?

MR. HYNEK: Twenty Feet.

MR. HOEFLE: They are twenty feet length, not twenty-six?

MR.HYNEK: No, that is right, twenty feet.

MR. GROVER: You indicated you have got a six stages unit and indicated that when nitrification starts up in the summer, so it takes awhile to develop that culture. Have they followed that culture, have they followed the build up of nitrites and then conversion of nitrites and get some quantitative estimate of the time period it takes to get nitrification?

MR. BERNER: No, not specifically at Gladstone.

MR. GOVER: And where does nitrification occur within that series of reactor system? Where do they see it picking up?

MR. BERNER: They see it picking up in the fifth and sixth stages.

MR.GOVER: And not prior to that?

MR. BERNER: No, not to my knowledge.

MR. STUDEE: Would you company be willing to provide an order with say a five year guarantee on structural failure of your unit?

MR. HYNEK: We have on a selective basis yes. As a matter of policy no we do not, unless everybody is required to provide a bond. We find ourselves decidedly disadvantaged at times, since we do not have five years experience with a multitude of failures we find ourselves penalized because we do not have that experience. And so a consequence we are many times required to provide a bond in lieu of the experience. On a selective experience we have. We would look at each individual job and decide that. As a matter of corporate policy we perfer not to. We do however we provide process warrenty.

MR. LUND: You have said that you are aerating the primary clarifier so as to enhance grease removal. What type of equipment do you have then ahead of the plant?

MR. BERNER: Grit removal is just a long rectangular channel with mechanical sludge collector. Actually our design of these primary clarifiers takes into consideration several things because of the infiltration in the sewer system, the clarifiers are oversized, what that means is though that in summer we had a low flow period of time like during July or August, we had a very long detention time in those clarifiers. In order to keep the clarifiers from going septic and that sort of thing, the aeration in the center feed-drive clarifier does a nice job of that. So we get a lot of benefits from aerating the primary clarifier, it is our favorite design.

MR. HYNEK: One minor correction. All of the consumption measurements you referred to in 1975 were done by an independent electrical engineer and subsequently these were found to be error. We had sponsored those initial prior measurements to get the field data and subsequently did this at other locations and found that the prior measurements were dramatically lower than those at Glastone. We then went back to Gladstone about two years later and made measurement again found they were about half of what the original measurements indicated. So rather than a consumption of approximately four horsepower per twenty-foot long contactor, it was really about two or two and a half horsepower per unit. So the total energy consumed by the six contactors is about twelve to fifteen horsepower.

MR. BERNER: I should appreciate it Mr. Hynek if you could send that data to me in the near future.

MR. MORGAN: You indicated the nitrogen removal problem was caused by pH. I was wondering did you test it in terms of alkalinity being the controlling factor?

MR. HYNEK: Well, the inflow alkalinity is approximately 150 mg/l and of course the alkalinity is a good indication of the nitrification capacity but it is not all. We have noticed that approximately 7 mg/l decline for each mg/l of ammonia removed. And so after the removal of ammonia in the first stage of the RBC, the alkalinity was reduced down to less than 100 which is the beginning of the subneutral pH. So I wonder if whether I have answered your question?

Session 10. NITRIFICATION AND DENITRIFICATION
Presiding: E. J. Opatken
Wastewater Research Division
U.S. Environmental Protection Agency

MR. ODGAARD: I am not going to question but I would like to show you a graph which is very similar to what was shown right now. These are the figures that I presented earlier in the conference. What I wanted to show is that the nitrification rate that was found in this study is exactly the same nitrification rate as I found in my study. Here is given as milligrams per square meter per hour. You gave yours in grams per square meter per day. When I calculated these I found four and that was the same as you counted. Another thing I wanted to show is that you showed that you got denitrification when you recycled. If you want to have of course good denitrification, you should use like this; you should use an anaerobic reactor before because then you can have the total denitrification as shown in this graph where the total nitrogen removal is shown versus the recirculation ratio. From that balance the possible nitrogen removal will be R divided by R+1. And these are the results of the study which fits this plot quite well.

MR. HYNEK: Are you doing any microbial study now with the RBC system?

DR. LONG: We have not. I guess the best way to answer the question is, part of our protocol identifies the kind of microbial work we anticipate doing. At the present time we have not specifically incorporated in that protocol into account, but it is anticipated that we will be looking at those in association with a nitrosonomas and nitrobacter. If the answer is yes we will be attempting to do that also.

MR. VESIO: Do you intend to try to use some high concentrations of ammonia and test those reactions on the fixed film?

DR. LONG: Our present thrust is towards domestic wastewaters, if we have enough funel, we may be making it higher concentrations.

DR. SAUNDERS: What kind of pH control system are you going to work with and what range or ranges of value are you going to investigate?

DR. LONG: Our effort on this study are a wide spectrum of pH initially in terms of acclamating the film. We are going to have from six all the way up to about nine, that is our initial efforts. Based upon what we see, in that first phase, we will then attempt to zero in on an optimum range or an optimum point. But we are going to wait until we see what the results are of our first phase and at that point make a judgment as to what we are going to be shooting for. In response to your first question, we have been looking at alternative chemical additions and possibly even aeration, various points of addition, perhaps even combinations. Again that will be determined a little bit later too.

DR. SAUNDERS: Are you going to control pH in each fill or just pH in the input?

DR. LONG: One of the configurations being envisioned now is to take a look at pH control coming into the influent, an alternative maybe also in stage three. We have now identified ts control pH control in one, two, three and four, that might not be too feasible. Right now we are anticipating, in anticipation of two schemes in terms of points of addition. We may find as we progress with our work that that does not quite do it and we might have to go to more points or perhaps just the first point might be satisfactory.

DR. MOLOF: The question of how much did these aeration or pH adjustment affect nitrogen balances in determining the activity of the biological oxidation and nitrification.

DR. LONG: We will attempt to identify the nitrogen balance that occurs and that loss which might be due to elevate pH in the stripping.

DR. HAUNG: Now what kind of medium do you use to control pH?

DR. LONG: There is no state of the method for identifying nitrosomonas and nitrobacteria. However the most commonly referenced media that is used is that of Dr. Alexander out at Cornell which he published in A.P. Black microbiology series about a decade ago. That is in essence both the media and the indicator that we are using. In addition to that we have made remodification to his media. We are adding a metal solution which he does not identify, but we feel is perhaps appropriate.

DR. YU: You are talking of adding addition of metals. What kind of metals do you use?

DR. LONG: I can provide you with that perhaps after. I do not have that list but there is about five metals in the solution that we add.

MR. ANDERSON: Would you differentiate between nitrite and nitrates? If you did, what test would you use?

DR. O'SHAUGHNESSY: We spot checked for nitrite and nitrate. There was never more than one tenth of a milligram per liter.

MR. ORWIN: I was wondering if you checked the cost effectiveness of increasing your volume of surface ratio. You said you got better reaction with everything but the greater volume. But that is going to cost you money. Do you know whether it is worthwhile doing?

DR. O'SHAUGHNESSY: Let me cite this just a minute before I answer your question. At the point four gallons per square foot of a rectangular reactor with the RBC, we put a small pump to keep everything in circulation. Now we are not trying to optimize the geometry. I do not even know if you can get it that high for a full scale unit. I think before you would want to go that far, you would try it out with some type of prototype, but you have to make sure your suspended solids will stay in there and not settle out. The cost effectiveness, just off the top of my head, you are talking about a little bit more concrete.

MR. ORWIN: You are actually talking about three times or three or four times the volume, so it could be fairly expensive.

DR. O'SHAUGHNESSY: Yes, that would be an engineering decision, right? I do not know what the largest one you can use is, no, we did not, I am sorry.

DR. SRINIVASARAGHAVAN: I have two questions. One is in connection with the slides that you showed. You increased the surface to volume ratio. The treatment efficiency increased but at the same time your loading values are also increased.

DR. O'SHAUGHNESSY: No, the loading values are based on pounds of ammonia nitrogen per thousand square feet per day, that is constant. Each unit has the same surface area.

DR. SRINIVASARGHAVAN: According to the loading numbers that you showed the two units were increased.

DR. O'SHAUGHNESSY: In the first part of the study, we had the Autotrol and the Hormell unit. They were pretty close in terms of surface to volume ratio; 0.09 and 0.12. Only in the spring portion when we drastically increased the surface volume ratio was that different.

DR. SRINIVASARGHAVAN: So, I wonder what the conclusion was so if you increased the loading the treatment efficiency increasing 15 to 20 to 35 or something like that. I was wondering if at the influence of the increasing volume to surface area ratio as you increased the ammonia influent concentration and therefore you got higher reduction.

DR. O'SHAUGHNESSY: No, we kept loading it at the same rate.

DR. SRINVASARGHVAN: Second question; what are the conclusions from the study that the staging does not influence the ammonia treatment efficiency?

DR. O'SHAUGHNESSY: No, we found that there was a definite impact of staging when you get fluctuations, you can see that. A lot of time indicated that the staging did make a big difference except in the two units. The fourth stage every time there was an upset because of temperature. Second, once we got onto the field we had much better performance as soon as we had two stages. So you would recommend a minimum of two stages than probably four or something like that. Yes, staging is something you would want.

MR. BELSCHNER: Did you measure suspended solids into an outer reactor and how did you determine any net sludge yield if any from the nitrification system?

DR. O'SHAUGHNESSY: Yes, we took effluent and influent and suspended solids at each stage.

MR. BELSCHNER: Was there a net sludge yield for nitrification?

DR. O'SHAUGHNESSY: We had just what was produced in terms of pounds of suspended solids versus pounds of ammonium removal.

MR. BELSCHNER: Was that based on solids coming out of the reactor or the difference between what went into it and came out?

DR. O'SHAUGHNESSY: The difference between the two.

MR. BELSCHNER: That looks unusually high. Typically we find a net solids destruction of secondary effluent solids going through the process when nitrifying.

DR. O'SHAUGHNESSY: We check the influent minus the effluent, plus, you know, what was there. And that is what it came out to be.

MR. BELSCHNER: That is an unusual result. With respect to that question, what would a solid concentration in the reactor systems, Jim? You show effluent concentrations higher than the reactor concentrations.

DR. O'SHAUGHNESSY: I may not have that information right here. They were not large, well under 50 milligrams per liter as I remember. We have the data. I do not know just what it is, I would say fifty. There was very little food in the system.

MR. BELSCHNER: Did you make the measurement of ammonia?

DR. O'SHAUGHNESSY: We measure ammonia using a specific electrode right at the plant.

MR. BELSCHNER: Is that an acceptable technique or...

DR. O'SHAUGHNESSY: Yes, we standardized the program in monitoring the system. Incidentally, I should point out how we will take samples. We took many samples based on hydraulic flow time. We would, if it was eight o' clock in the morning and we knew it was one hour detention time we would take the next one at nine, ten, eleven, and so forth and follow that plug through. And if they came out not close enough we would not take the third, and results are based on average value.

MR. BELSCHNER: When you are measuring efficiency of the system, you are talking about efficiency of conversion or removal of ammonia under dynamic loading conditions as well as organic loading in the system. It would seem like you are attributing to nitrification a certain amount of conversion, a variable amount of conversion of nitrogen to cell mass in the system. Won't you maybe look at nitrates as a by-product to determine the process efficiency? One figure I thought was bad, maybe I misinterpreted.

DR. O'SHAGHNESSY: No, I am sure, the effluent nitrate concentration was an average from the study; both studies, 90% of the influent ammonia concentration, so 90% came out in our effluent as nitrate, 10% went into solid mass or denitrification or whatever else you want to attribute it.

DR. ODEGAARD: I would like to know how do you plot your results, because it was not obvious that any maximum nitrification rate, and if you are going to design a nitrification system we have to know a little bit about the kinetics of the nitrification. So my question is, did you find any when you increased your loads; did you find any plateau where you had a biological reaction? Did you go into each of the steps and try to find a Monod relation?

DR. O'SHAUGHNESSY: No, as I tried to point out in the beginning, this was intiated in 1976. One of the concerns is can we achieve and what do we need to achieve one milligram per liter of a secondary effluent, one milligram of ammonia in these wastewaters. We found this was what we needed to go up to get equilibrium to find out what would be the limiting factor. You would be well above the loading rates obviously as the other information has been presented here. And you would be close to one milligram per liter. And this was not that big a study, so we did not look at what you are interested in.

DR. ODEGAARD: I mentioned this because I think it is interesting. If you go into the literature and we heard from the Japanese presentation today and my presentation, our figure and all the figures that I found in the literature are very similar. So the maximum nitrification rate is similar. Then of course, we have an obvious difference in conclusion.

DR. O'SHAUGHNESSY: All well, On the paper presented this morning, I went to calculate just roughly. I am estimating that you know, loading about ten pounds of ammonium nitrogen per thousand square feet per day, and the highest level we got was point eight pounds.

DR. ODEGAARD: I have a problem converting.

DR. O'SHAUGHNESSY: I think if you look at that I think you will find its value almost identical to us or two off in terms of where I am. We are interested in what is going to happen to achieve less than one milligram per liter as required by a discharge permit.

MR. MOLOF: You have four stages RBC system. What are the observations in terms of nitrification versus temperature in each stage? Did you make this study?

DR. ZENZ: We did and I did not present that information because of time consideration. During the test period where you saw, when we showed information on 80 percent, we were getting nitrification in the last stages. Now some of those higher removals with the higher temperatures, we were not getting a lot of nitrification in the very last stage. So it would shift; it would generally depend on temperature, but in the numbers that I am giving you winter time temperatures we were getting 80% removal. The entire unit is nitrified and it is you know, I am not giving you information whether it is unneeded stages at that the matter of the question you gave me.

MR. COULTER: In your last slide you had 80% removals at cold temperatures at both point two and point four loadings. This is a substantial difference in terms of hardware required. Did you have any supplementary data in terms of alkalinity or anything else that might explain that difference?

DR. ZENZ: No, I cannot really explain it. There were changes in influent ammonia concentration, unfortunately I cannot remember if those went with those particular places. There was some scatterings in the data. I cannot really tell you right now if I could postulate a reason for that off the top of my head.

MR. WATT: Do you have any leakage of ammonia through your nitrification system?

DR. ODEGAARD: No it was quite low, It was totally nitrified.

MR. ATHAVALEY: Since we are talking here about anaerobic denitrification what is the temperature factor in removing the ammonia nitrogen? Did you study the temperature effect?

DR. ODEGAARD: No we did not. As I said we did not even have a temperature controlled, but the temperature in these experiments were between 12 and 16 degrees centigrade. But you would normally not use heating of anaerobic step anyway. So I do not think it is...of course the temperature influence is interesting to all of us.

MR. HAUNG: Did you use any organic compound in denitrification step to achieve 1mg/l of ammonia nitrogen in the process effluent?

DR. ODEGAARD: You mean to take out the rest of ammonia? No, I think there is a plant in Japan which is similar to this one, but in addition, have a second step where they use methane to take out the last part.

DR. HAUNG: It appears that in a tremendous hydraulic loading to the denitrifier, and I wonder would there be a minimum retention time upon this application. Would that affect your results?

DR. ODEGAARD: That is an interesting question because we worried about that. But the only problem we saw is that when you operate the recirculation to very, very high levels we tried to go as high as twenty, then we got problems. As long as we stayed under ten and the plant is designed of the normal criteria. so that the retention time as such is short, then we did not have any problems.

DR. HAUNG: What is the retention time?

DR. ODEGAARD: That is a good question. I think, I do not have it in here right now but it is, well, that depends upon what you are asking about because the theoretical detention time through the plant is I guess approximately three hours.

MR. BALANCE: What was the retention time meant in your case?

DR. ODEGAARD: Well, if you look at one particle, how long the particle is retained in the system. But then the water was circulated all the time so it comes back, the same particle comes back many times of course. So that how do you define detention time?

Session 11. SELECTIONS AND ECONOMICS
Presiding: J. Miller
Department of Civil Engineering
University of Pittsburgh

MR. JOOST: Keep this in your mind, the power consumption for the mechanical drive is quite high to me.

MR. BARRY: I have heard numbers anywhere from three to three and a half to four. That has been down since we did this analysis. So again, it is going to be an analysis like this has to be done on a case by case basis on all rates and the best information on it to find.

MR. BERNER: Can you comment on your solid handling processes at this plant?

MR. BARRY: The solid handling for RBC sludges are going off the landfill immediately. The primary sludge would be pumped directly to anaerobic digester. The secondary sludge there are provisions for it to go to either a gravity thickener with or without chemical polymer or directly to the digesters. The two stages digesting system would be recovering the gas and running into generators from that system after the solids have digested they will be taken off the plant site by tanks, trucks, and disposed of on the farmfields. We found the farmers are very receptive to this humus waste material because of the nature of the waste product. And anything the farmers can get that will hold moisture in the soil they are very anxious to get.

DR. SACK: How did you handle labor costs, number of operators, things like that?

MR. PIERCE: Well, actually the number of operators is somewhat debatable between analysts. We felt that in general the RBC processes should have fewer operators and if there was any differentiation between the two the mechanical drive RBC installations, particularly the large ones would have perhaps an additional one operator. In general we only added for the activated sludge processes sufficient operators to handle the thickening aspects. The estimation was the largest design plus three more people operating an activated sludge plant than operating RBC plant.

MR. BAO: Do you have any comments about Dr. Sun's conclusion about what he said, his recommendations he said the activated sludge plant would be cheaper than RBC.

MR. LUNDBERG: I do no recall his paper. He is not here to defend himself on any comments. I think he was talking about synthetic trickling filters versus the RBC in that application. I do not think he discussed activated sludge in his paper and somebody can correct me if I am wrong on that. I am expecting this question was going to arise. I really do not necessarily see a conflict there because we are talking about three separate applications where Pierce was talking about carbon oxidation alone. I was talking about combined carbon oxidation and nitrification in a single stage and he was talking about simply nitrification alone in these reactors as a simple process onto itself. So to begin with we were not talking about the same thing. The other thing that crossed my mind was the design criteria that was utilized in his paper versus mine. Now it was two different applications. I think that if we tried to sit down and compare apples to apples, that he was maybe using less conservative loading rates for the synthetic media filters than I did. Now I am not sure so as I said, first of all I really do not see it as a bumping heads type of thing, because we thought it fanned out rather nicely that there were three papers here that talked about three separate types of applications where Pierce and I had two of them and we opened up the list of people to the symposium that goes with two-thirds of the paper complement that we were talking about. Those are the only two things I can see we have to look at specifically at the cost factors he used and the design criteria to really figure out what the story was.

MR. GROVER: The last speaker, did you make any assumption in the different sludge dewatering characteristics that purported to exist within the pure oxygen design? Would you assume a better dewatering characteristic maybe due to chemicals addition?

MR. PIERCE: Well, actually we did not get into the handling aspects. In this study it was basically liquid treatment and thickening. We did assume a higher underflow concentration in the pure oxygen case of perhaps one point seven five percent underflow versus point seven five percent underflow for the air, and we did include a difference in the dissolved air flotation loading rate. Our opinion was that regardless of the sludge handling train utilized, the liquid treatment of thickening could pretty much be separated and any...there is a great deal of debate about how much sludge one gets from an RBC process and whether it is or is not as dewaterable. I think in the worst case against RBC's they would all be pretty much the same. So the sludge handling could be addressed separately in a facilities plant.

MR. SUGAR: I guess I am going to have trouble with all of your curves shown previously. I do not know if you have realized that in particular, the curves you show with the trial versus the PSA for the pure oxygen plant. I guess I am surprised that there is such a similar delta that you are talking about, because our internal analysis we were trying to find whether there should be a PSA or a trial. You definitely tend to go with the PSA plant from the cost effective point of view, attempting power in the smaller range plant. And we have obtained a trial plant very much below twenty tons a day. On the other side of the coin, you really would not consider using a PSA. That would be an excessive to that and therefore I am somewhat surprised at that delta with that range process range.

MR. PIERCE: Well actually there are probably two explanations for this. It is a secondary treatment application and it has a relatively dilute low concentration of BOD. We did basically center this study around a similar study done at a thirty MGD range and our oxygen generation requirements were on the order of about eighteen tons at thirty MGD. That particular analysis had indicated the PSA had a slightly lower present worth value and being concerned about the small delta on here, it is really not that small if you are only considering the oxygen generation facility or the aeration tanks. But bear in mind that there is a large block added to all of those and the thickening facility, the final clarifiers and all the other unit process equipment for the whole plant if you were to take those, the differential and apply that entirely to the difference in oxygen generation facilities, which it pretty much is you would probably find that the delta is much more what you think it might be. I cannot really explain why in the range it fits the...it was still a PSA plant and I do not think that again it was in plus or minus ten percent. It could have gone either way.

MR. ANTONIE: I would like to have a point clarified regarding a question that was asked earlier, Mr. Pierce, regarding, in fact I do not believe the consulting engineers are still here...He asked what the impact in your conclusions would be, shorter evaluation period the sort of payback period, and he specifically asked that he was concerned with industrial wastewater applications. I would like to point out, you can correct me if I am wrong Mr. Pierce, that your evaluation was based solely on the kinetics of municipal wastewater and that you cannot extrapolate your report on industrial waste.

MR. PIERCE: I do not know if that was his specific intent but what you are saying is true. We assumed a typical municipal primary effluent, so depending on the fraction of soluble versus insoluble BOD in waste there is such a variety of other factors that the only way you could interpolate you know, off of this graph and say that for one MGD industrial installation many things apply here...I think he may just be giving that as an illustration of the fact of what would happen if somebody placed more budget upon capital costs than on the operating costs. But as you did say, I think there is a lot of danger in trying to take things somebody else did, particularly cost curves, and apply it to your particular situation. And I think Lee and I both feel that nobody should take the cost curves that we developed and take a look at them and say okay, if it is a 30 MGD plant, this is the construction cost. It has to be done on a case by case basis. There are just so many variables as far as we are concerned, and you can correct me if I am wrong. One of the cost curves we have developed is totally useless except to give you a general indication. I think what Lee wants to try to show was that the RBC process should not be reasonably thrown out the window as many people have done because we do think that the cost for a number of applications brought across a wide range of flow.

☆ U.S. Government Printing Office: 1980 M-654-177.